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ENFOR Project C-14

Oil Replacement on a Pulp Mill Lime Kiln Using Hog Fuel in a Lamb Wet-Cell Burner

by

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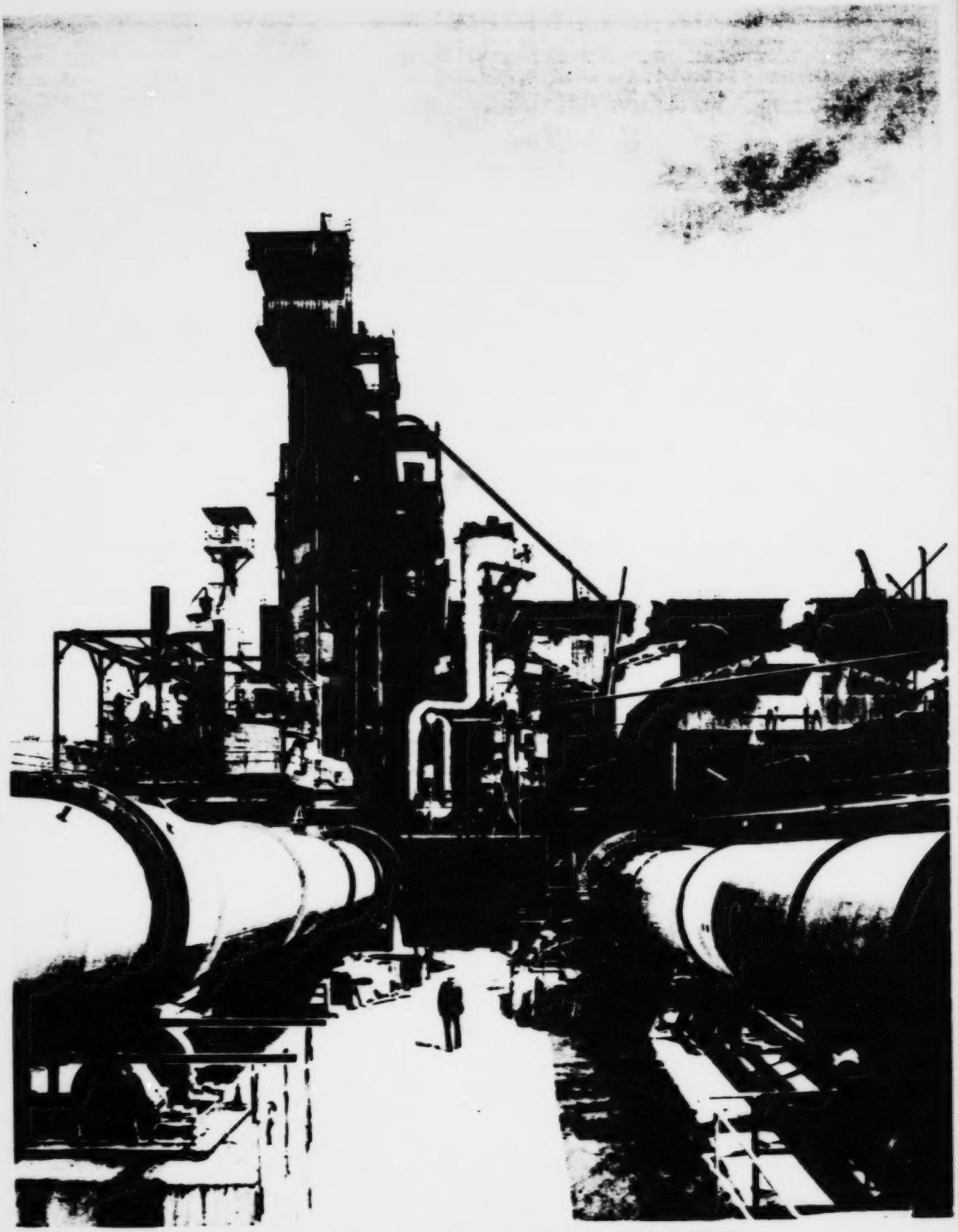

LAMB WET-CELL/LIME KILN CONVERSION INSTALLATION

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EXECUTIVE SUMMARY

OIL REPLACEMENT ON A PULP MILL LIME KILN USING HOG FUEL IN A LAMB WET-CELL BURNER

Background

Fossil fuel consumption in North American pulp mill lime kilns is currently estimated to be 25-million barrels of oil (equivalent) per year.

Recognizing the enormous savings which would accrue from conversion to wood waste fuel, a number of Canadian Government agencies cooperated with industry in funding a demonstration project.

Objectives

The objectives of the project were as follows:

(a) Design, installation and testing of the Wet-Cell Burner and kiln conversion system.
(b) Identification and implementation of the necessary technical modifications to permit dual-fuel operation.
(c) Evaluation of the performance limits for dual-fuel operation.
(d) Use of computer simulation techniques to assist evaluation of the performance limits and to permit extrapolation of the data for other kiln conversions.

The report discusses design, thermal and economic aspects of simultaneously firing a lime kiln with oil and hogged wood waste combustion products.

System

The test kiln, located at the MacMillan Bloedel Kraft Mill, Port Alberni, B.C., Canada, was 7' dia. x 200' long with a design rating of 53 tonnes/day lime product.

The Wet-Cell Burner, rated at 63 GJ/h (60-MM Btu/h) complete with fuel system, automatic ash removal and hot gas ducting, was installed without interruption to mill production.

Conversion of the kiln necessitated the following modifications:

provision of a hot gas inlet breaching on the firing hood;
increased induced draft fan capacity for larger gas flow;
provision of a compatible control system for dual-firing;
air cooling of the oil gun assembly and kiln nose ring; and
placement of refractory in the kiln hood to minimize surface metal temperatures and to redirect Wet-Cell gases inside the kiln.

Typical fuel for the tests was coastal wood waste at 55% moisture content, hogged to pass 75 mm grid (3" minus) with a higher heat value of 20.97 GJ/dry kg (9018 Btu/lb). Typical gas temperature at the Wet-Cell Burner outlet on this fuel was 1200° C. (2200° F.).

Trials and Results

Representative baseline test data were recorded for oil-fired operation in order to establish the impact of dual-firing. Baseline lime production on test was 37.3 tonnes/day. This corresponded closely to normal mill operation but was below the design rating due to mechanical feed system limitations on this kiln. The thermal analysis of the calcination process was unaffected by this constraint. Dual-fuel tests were conducted with volumetric mud feed to the filter held constant at baseline level. Test data acquired reflected the impact on kiln performance of heat input, gas temperatures, ingress air, percent solids in the mud feed and dust loss.

Oil substitution is defined as the percentage reduction of baseline oil flow and replacement ratio as the number of wood-derived thermal

units used to replace each oil-derived thermal unit at baseline.

For several levels of substitution, wood-heat input was varied with fixed oil flow to determine the effect on lime availability. Lime availabilities were maintained at substitution levels up to 78% with dual-firing.

Replacement ratios were greater than unity, and hence energy input/tonne product higher than for 100% oil-firing at baseline.

The fundamental cause of reduced efficiency on wood heat was the lower heat transfer resulting from reduced gas temperatures. This also resulted in high exit gas temperatures which were restrained by additional water on the mud screw. Increase in total gas flow to compensate for lower gas temperatures also created higher losses, and cooling air at the hood and oil gun contributed to reduced efficiency, but the latter can be minimized in conversion design.

Computer Simulation

Boeing Engineering & Construction Company was retained to provide the lime kiln simulation and a continuous data acquisition system for the project. The simulation had been developed and used to optimize numerous oil and natural gas-fired kilns. Based on the extensive data from the dual-fired kiln tests, the simulation required adaptation to accomodate the increased significance of convective heat transfer and the gas radial temperature profiles.

The simulation was used to develop parameter profiles providing an insight into the thermal aspects of dual-firing for the test kiln. This work was then extended to predict optimal performance of the test system and evaluation of performance limits for dual-fuel operation.

Prediction of Optimal Performance

The simulation indicated that efficient conversion of an existing kiln should include consideration of the following:

(a) Reduced kiln speed
(b) Increased refractory thickness
(c) Increased Wet-Cell gas temperature (lower moisture content fuel)
(d) Reduced ingress air.

The influence of these factors is discussed in the report. Feasibility of such improvements will vary according to engineering and economic factors in each specific case.

Optimization of the test kiln was explored assuming reduction in speed and ingress air only. Kiln diameter and length were unaltered and Wet-Cell gas temperature was restricted to the 1200° C. value achieved on test with fuel at 55% moisture content.

The optimized performance predicted by the simulation indicated replacement ratios of 1.18 - 2.28 at substitution levels of 50 - 80% respectively.

Test values of replacement ratio at similar levels of substitution were generally greater than 2.0 It is recommended that a computer simulation be employed for evaluation of dual-fuel conversions on other kilns.

Conversion Design

In approaching conversion particular attention should be given to the following areas:

Heat Source: Typical fuel at the test kiln location was 55% M.C. Where the available M.C. is below 50%, increased gas temperature can be achieved from the Wet-Cell Burner with benefits to kiln performance.

Hot Gas Duct: Radiant heat loss should be minimized through generous refractory thickness and shortest possible duct length.

Kiln: Economic limits of induced draft fan capacity and reduced kiln speed should be exploited. Scrubber improvements may be required to minimize dust loss.

Kiln Hood: The conversion of the test kiln was accomplished with minimum hood rework. The tests and simulation suggest objectives for hood redesign should include improved nose ring and oil gun cooling; reduction of ingress air and reduced turbulence at hot gas entry. These features are incorporated in a proposed hood design.

Economic Analysis

Decisions on kiln conversion are dictated by a combination of engineering and economic factors. The latter were investigated for a lime kiln rated at 100-MM Btu/h on No. 6 fuel oil, displaced by 55% M.C. hog fuel, with the following conclusions:

1. Based on current world residual oil price, lime kiln conversion is economic, if hog fuel is available for $30/unit or less.

2. The optimal level of oil substitution will vary with the prices of oil and wood waste, the capital cost allowance scheme in effect, and the moisture content of the wood

waste. For low oil or high hog fuel prices, a low substitution level (50%) is preferred. For high oil prices or low hog fuel prices, a high level (75%) is best. For intermediate fuel prices, an intermediate level of substitution is preferable.

3. The introduction of a two-year tax write-off substantially improves the project's profitability and has a greater effect on profits for projects replacing more oil. Shifting from straight-line depreciation to a two-year write-off will, in some instances, make higher substitution levels preferable.

1. INTRODUCTION

The production of lime (CaO) from calcium carbonate mud or limestone is an essential part of the chemical recovery cycle of the Kraft pulp process. The calcination reaction:

$$CaCO_3 \longrightarrow CaO + CO_2 \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (1)$$

is endothermic with a heat of reaction of 1.69 MJ/kg (730 Btu/lb) $CaCO_3$ at 900° C. (1650° F.). To provide this energy, the majority of mills use rotary oil or gas-fired kilns. The pulp and paper industry in B.C. alone is using gas and oil equivalent to 7,600,000 barrels of oil per year (1). Typically, a coastal bleached kraft mill will consume 44% of total oil in the firing of the lime kiln.

TABLE 1 TYPICAL FUEL CONSUMPTION OF A B.C. COASTAL MILL PRODUCING 600 METRIC TONS/DAY OF BLEACHED KRAFT PULP (2)

	Total Fuel Use (Bbl oil/yr) equivalent	Energy From Liquor Recovery (Bbl oil/yr) equivalent	Energy from Hog Fuel (Bbl oil/yr) equivalent	Energy from Oil (Bbl/yr)	% Energy from Oil
Lime Kiln	103,500	0	0	103,500	44%
Recovery Boiler	445,050	414,000	0	31,050	12%
Hog Fuel Power Boiler	414,000	0	310,400	103,500	44%

At an annual Canadian production capacity of 7.3-million tonnes (8.0-million tons) kraft pulp, the minimum fossil fuel requirements of the industry for lime kilns may be estimated at 3.3-million barrels per year.

For the pulp mill lime kilns operating in Canada and the U.S.A. alone, total consumption is estimated at 25-million barrels oil equivalent per year. The potential for North American and world export of a fossil-fuel replacement technology is therefore considerable.

The Lamb Wet-Cell Burner, developed by Lamb-Cargate Industries Ltd., of New Westminster, B.C., is capable of generating clean, high-temperature gases (1260°C., 2300° F.), using green wood or bark as fuel. Since no data was available to predict kiln performance using wood fuel, Boeing Energy Control Systems was hired in August, 1977 to run a computer simulation of the Lamb Burner firing a lime kiln. This study indicated that 60% of the heat load might be provided by the Burner without derating an existing kiln.

In 1978, agreements were drawn up by Lamb-Cargate Industries Ltd., MacMillan Bloedel Limited, the Canadian Forestry Service ENFOR Program and the Canadian Department of Industry, Trade and Commerce Economic Development Program to enable a Lamb Wet-Cell Burner System to be designed, manufactured, installed and tested as a heat source on MacMillan Bloedel's No. 1 Lime Kiln at Port Alberni, British Columbia. Later, additional funding was provided by the British Columbia Ministry of Energy, Mines and Petroleum Resources to complete this phase of the work. Following encouraging results from initial operations, the National Research Council of Canada provided funding to cover the evaluation of the system with modifications and testing to optimize wood-waste firing of the lime kiln.

Calcination Process

Calcium oxide from the kiln is reacted with green liquor ($H_2O + Na_2CO_3 + Na_2S$) in the slaker to form calcium hydroxide ($Ca(OH)_2$). The reaction proceeds to the causticizer and

clarifier where calcium carbonate sludge is precipitated and white liquor ($NaOH + Na_2S$) is produced for the pulp digesters. The calcium carbonate sludge is returned to the kilns via mud washers and filters. Limestone or lime is added to the cycle as required on a make-up basis. The process is illustrated in Fig. 1.

The rotary kiln is a horizontal counter-flow reaction vessel and its efficiency may be measured in terms of its specific energy consumption. Oil-fired kilns may use 9.3 to 15 GJ/tonne CaO (8 to 13-million Btu/ton CaO) produced, depending on operating variables such as feed rate, rotational speed, inclination angle, combustion efficiency, feed moisture content, refractory condition, etc., and upon use of auxiliary equipment such as chains for mud drying, air preheaters or recuperators, and oxygen-enriched firing systems. In addition, computer control systems can improve user efficiency.

Lime Kiln Conversion -- State of the Art

The typical rotary lime kiln is designed specifically for oil or gas-firing. It is probable that different types of reaction vessels would be proposed for the design of new installations with 100% wood waste firing. The successful conversion of existing systems depends on an understanding of the physical limitations of the equipment and of the heat transfer process involved. Specifically, good combustion of oil will produce adiabatic flame temperatures around 2000° C. (3600° F.). In contrast, dry pulverized wood combustion in suspension burners may generate 1450° C. (2650° F.), while a well-designed hog fuel furnace can achieve gas temperatures of 1260° C. (2300° F.).

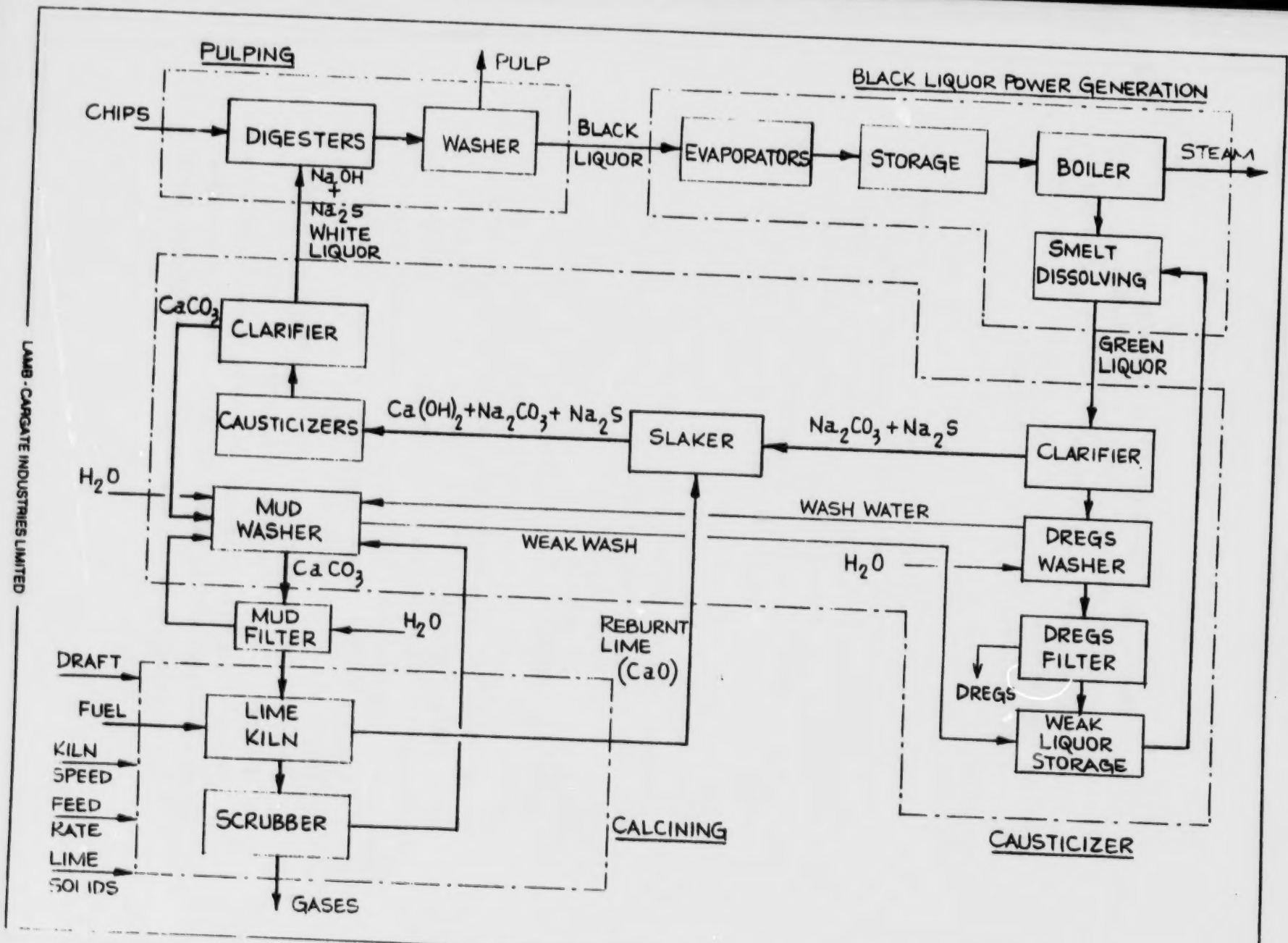


FIG. 1 CHEMICAL RECOVERY PROCESSES IN A KRAFT PULP MILL

Wood-fired conversions must accomodate reduced thermal gradients and their impact on heat transfer. Two systems are described below.

The Flakt Bio-Fuel System

This system was installed at an ASSI Kraft Mill in Sweden. Two kilns and one boiler are fired with dual-fuel, suspension type cyclone burners.

Fuel is dried and pulverized to 0.5 to 0.15 mm in a complex and costly preparation system. The primary air is not heated. Secondary air is heated by product lime-coolers. The system gives actual flame temperatures of 1450° C. (2650° F.) and has achieved high or total oil substitution on these kilns.

The apparent disadvantages of this system include the high capital cost and energy load imposed by pulverizing and drying the fuel, and the impact of contaminants (sand or ash) in the combustion products. Fuel preparation costs have been justified by increased thermal efficiency of the boiler, and contaminants are reported to be removed in the bowl classifier and white liquor clarifier.

In common with all wood-fired systems, the fuel is sulphur-free and sulphate or sulphite levels are substantially reduced compared to oil-firing (3).

The system is commercially available in North America.

Wood Residues as a Fuel Source for Lime Kilns
(ENFOR Project C-123)

This project explored the advantages of feeding hog fuel into the kiln with lime rock or mud. Combustion was initiated with oil or gas, becoming ultimately self-sustaining. If proven, such a system would minimize fuel preparation, and would provide close thermal contact between feed and the fuel. The following is abstracted from a 1982 Report (4) on the Project:

> "Despite flow problems encountered when feeding wet hog fuel with mill-produced lime mud directly into the cold end of a pilot kiln, a decrease in fossil fuel consumption of approximately 20%, based on long-term average product flow rates, was obtained. The flow problems consisted of material adherence onto the kiln wall, increased dusting and feed-end losses.
>
> The lime produced with hog fuel showed higher reactivity than lime produced without hog fuel. Upon causticizing a high-sulphidity mill green liquor, the equilibrium causticizing efficiency was not affected by the use of wood residues in calcining. However, there was some deterioration in the settling and filtration characteristics of the lime mud due to the presence of hog fuel contaminants.
>
> The normal fuel requirement per ton of lime in mill kilns is only about one third of that in the pilot kiln. Because most of the problems and limitations encountered in the pilot kiln were associated with the very high

mass ratio of hog fuel to lime mud required to achieve even 20% replacement, fossil fuel substitution of 30 to 50% might be attainable in commercial scale lime kilns.

The ultimate goal of complete or near-complete substitution of fossil fuel by wood residues can be achieved by additional firing of dried (moisture content 10-15%) fines from the hot end of the kiln. Since this latter type of fuel is substantially more costly than wet, non-comminuted hog fuel, its usage should be minimized."

The above approaches utilized pulverized wood fines and direct mixing of hog fuel with lime. The Port Alberni project was designed to explore the feasibility of conversion using hog fuel combustion outside the kiln.

2. OBJECTIVES OF THE WET-CELL/LIME KILN PROJECT

In the wood conversion systems described in Section 1, major problems are associated with combustion within the kiln. These shortcomings (fuel drying, particle size reduction and product contamination) are accepted in order to achieve high temperatures or enhanced heat transfer.

The Wet-Cell approach is to ensure complete combustion outside the kiln, and thereafter to calcine with hot combustion products. The unit is designed to burn a wide range of hog fuel in the 75 mm (3") minus size range, at high moisture contents, with low particulate emissions, less than 0.1 grains/sdcf. Product gas temperatures are somewhat lower, typically 1260° C. (2300° F.) with 55% fuel moisture content, but still substantially above calcining temperature (900° C/ 1650° F.)

Major advantages of this approach include the use of proven burner technology and the elimination of high capital investment in fuel preparation and drying equipment.

The objectives of the project were as follows:

(a) Design, installation and testing of the Wet-Cell Burner and kiln conversion system.
(b) Identification and implementation of the necessary technical modifications to permit dual-fuel operation.
(c) Evaluation of the performance limits for dual-fuel operation.
(d) Use of computer simulation techniques to assist evaluation of the performance limits and to permit extrapolation of the data to commercial operations.

The availability of the kiln was a unique situation, in that it was only intermittently required for mill production. It was soon to be decommissioned due to age and mechanical problems, including

limited feed-rate capability. This limitation however, did not affect the acquisition of a data-base, from which an understanding of the dual-fired process was developed.

3. SYSTEM

Lamb Wet-Cell (Fig. 2 & 3)

The Lamb Wet-Cell Burner consists of two chambers, one above the other. The lower chamber dries -- then gasifies -- the wet wood waste. The volatiles thus liberated are burned completely in the upper combustion chamber creating hot, clean gas for the process. Efficient application of combustion air in both chambers ensures low excess air levels and therefore high exit temperatures in the range 925-1260° C. (1700-2300° F.).

The burner is designed to combust coarse, wet hog fuel (less than 75 mm (3"), up to 60% moisture content (wet basis), and generate a clean, hot gas at a variable controlled rate. It is capable of low excess air operation for high off-gas temperatures but avoids the associated problem of clinker.

The Lamb Wet-Cell Burner uses an underfeed stoker to inject hog fuel through the burner base, where it spreads over the pinhole grates to form a slightly conical pile. This minimizes tramp air ingress associated with moving grate stokers, and avoids particle carryover associated with gravity feeding. Specific details of the Port Alberni system were as follows:

Fuel and Ash System

Front end loaders drive up a ramp and dump hog fuel into the metering bin (Fig. 4). The bin has a capacity of 24 m^3 (850 cu ft) and can hold approximately two loads of fuel. At rated capacity (63 GJ/h) the bin requires filling every 20 minutes (55% M.C. fuel), (Figs. 5 and 6).

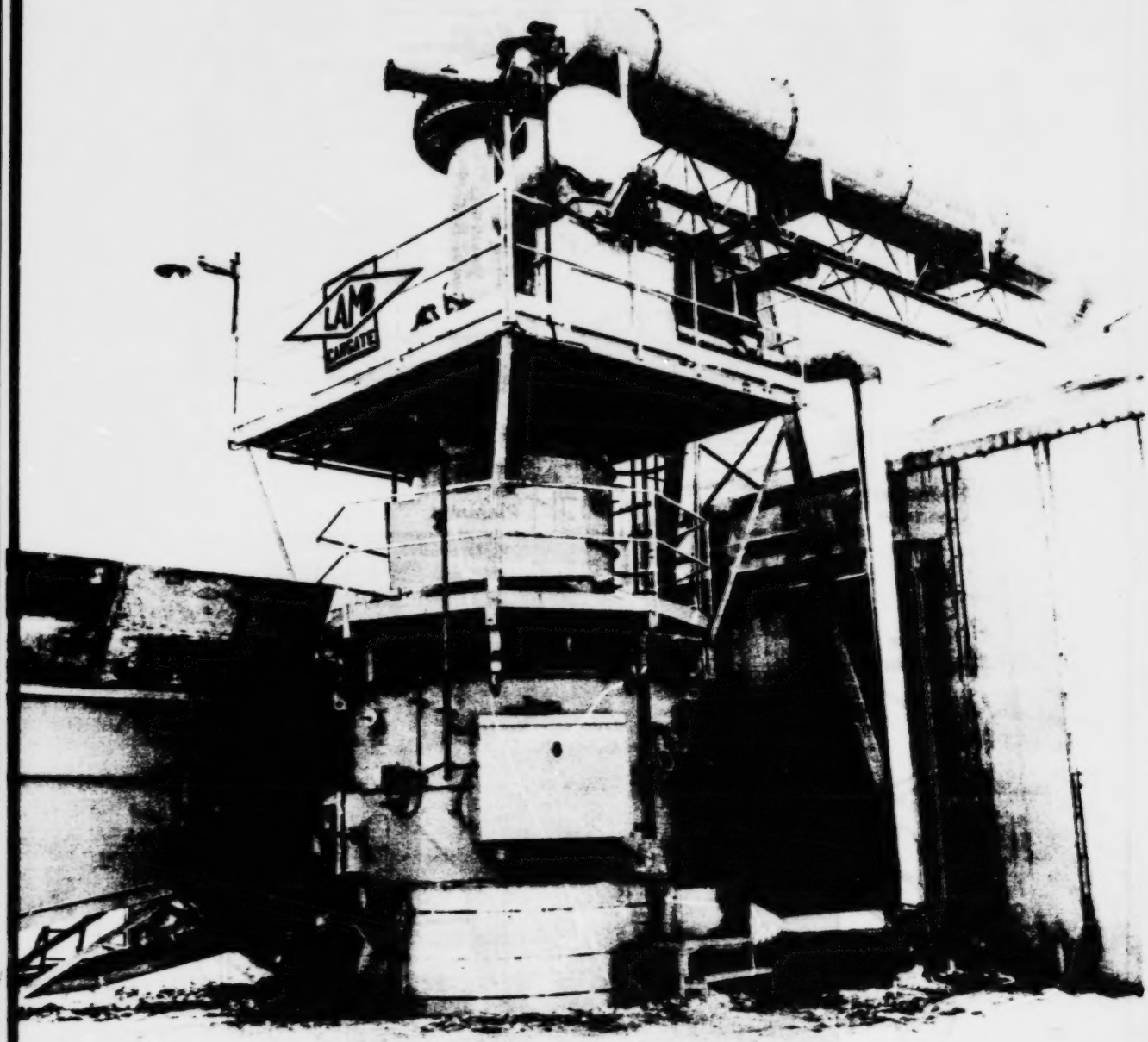


FIG. (2) LAMB WET-CELL

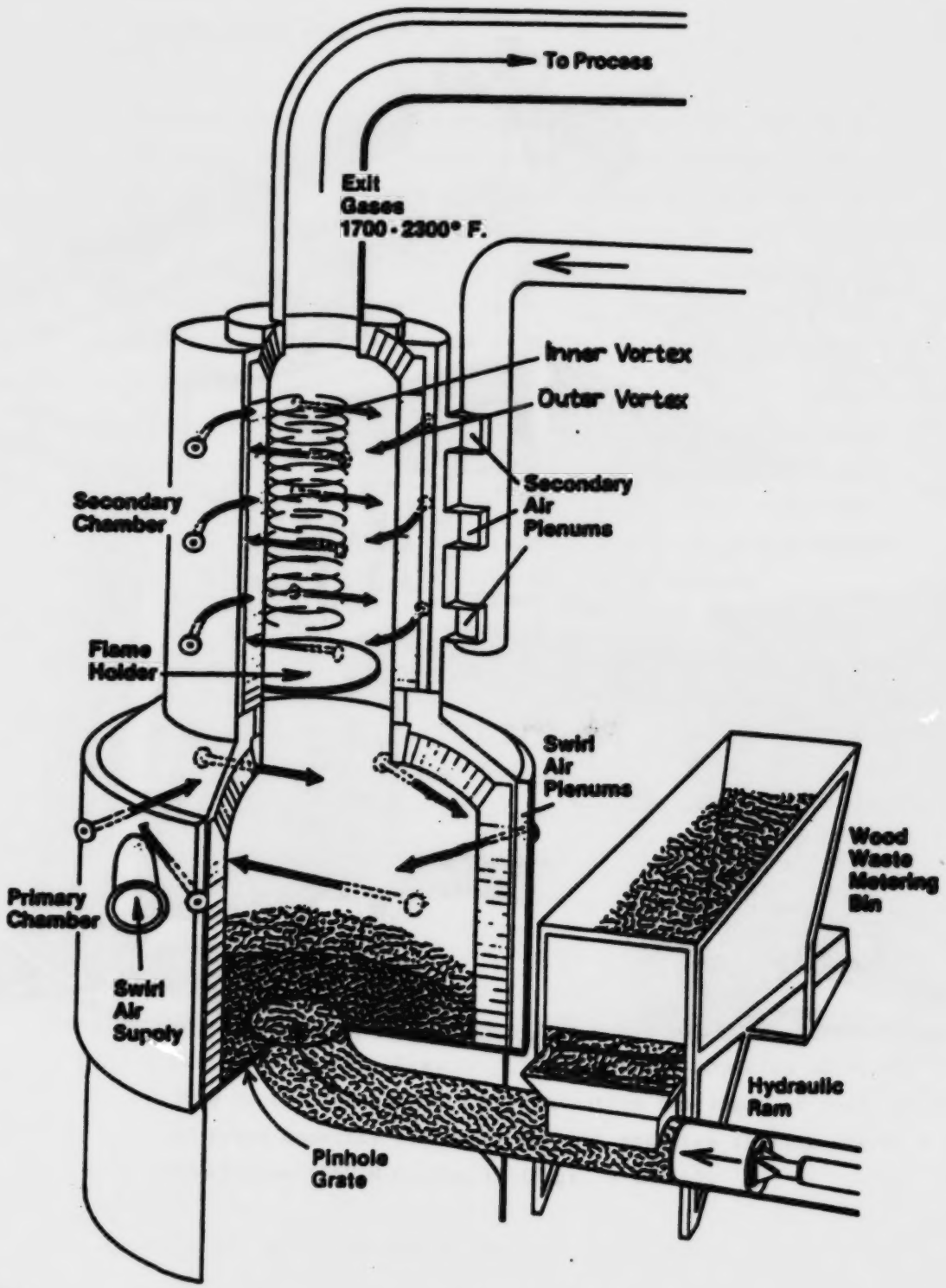


FIG. (3) LAMB WET-CELL -- SCHEMATIC OF PRINCIPAL ELEMENTS

FIG. (4) METERING BIN RAMP

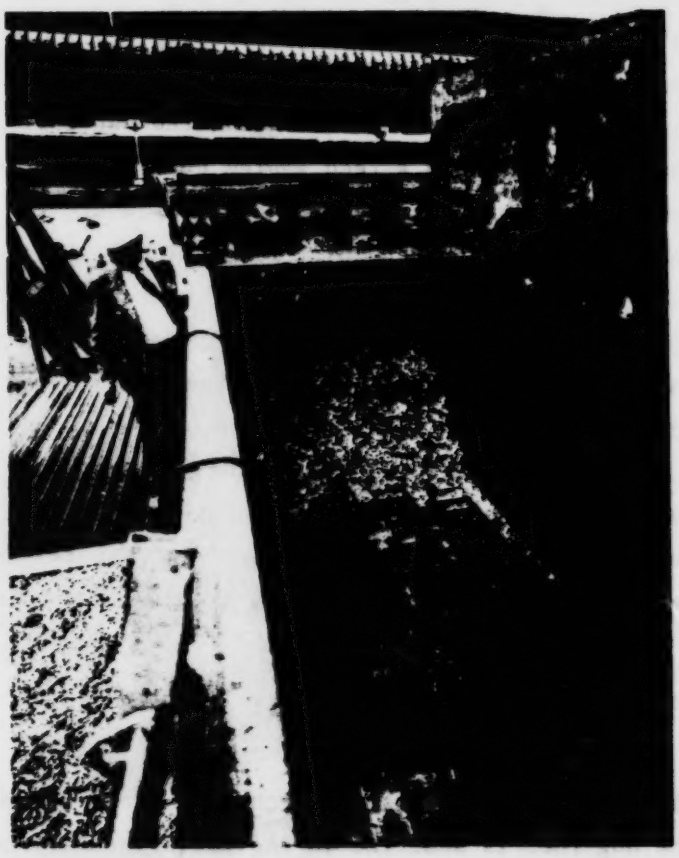


FIG. (5) METERING BIN -- INSIDE VIEW

The bin is provided with a live-bottom conveyor and a rake-back conveyor to prevent hog fuel jamming against the end of the bin and to ensure a regulated supply of fuel to the blowline for control purposes. The bin discharges to a rotating disc screen (Fig. 7) which removes oversize particles to an "overs" conveyor. The screen discharges to a belt conveyor equipped with a magnet for metal removal. The fuel then drops into a rotary blowline feeder/airlock.

The metering bin live-bottom conveyor and rake-back conveyor are driven by hydraulic motors. One hydraulic tank and two pumps provide the hydraulic power required to drive the total metering bin system with the capability of varying the speed according to the burner hog fuel demand. A 20 mm (8") high-pressure blowline, approximately 200 m (660 ft) long, pneumatically conveys the hog fuel to a 1524 mm (60") diameter cyclone mounted alongside the Wet-Cell Burner. The hog fuel drops into a feed system consisting of a horizontal and vertical screw conveyor, located beneath the combustion chambers of the burner. The vertical screw distributes the hog fuel in a fluffy condition on the central sloping portion of the grate. As combustion proceeds, ash is formed and removed automatically by the continuous ash discharge system (CAD). Ash gravitates from the sloping grate onto a refractory-covered rotating ring. A stationary plough pulls the ash off the rotating ring into an ash discharge chute equipped with two automatic slide gates so that a seal is provided when the ashes are dumped into an ash box.

Combustion and Hot Gas Delivery

The Wet-Cell Burner is a two-chamber incinerator. (Fig. 3). The hog fuel is admitted in the lower (primary) chamber,

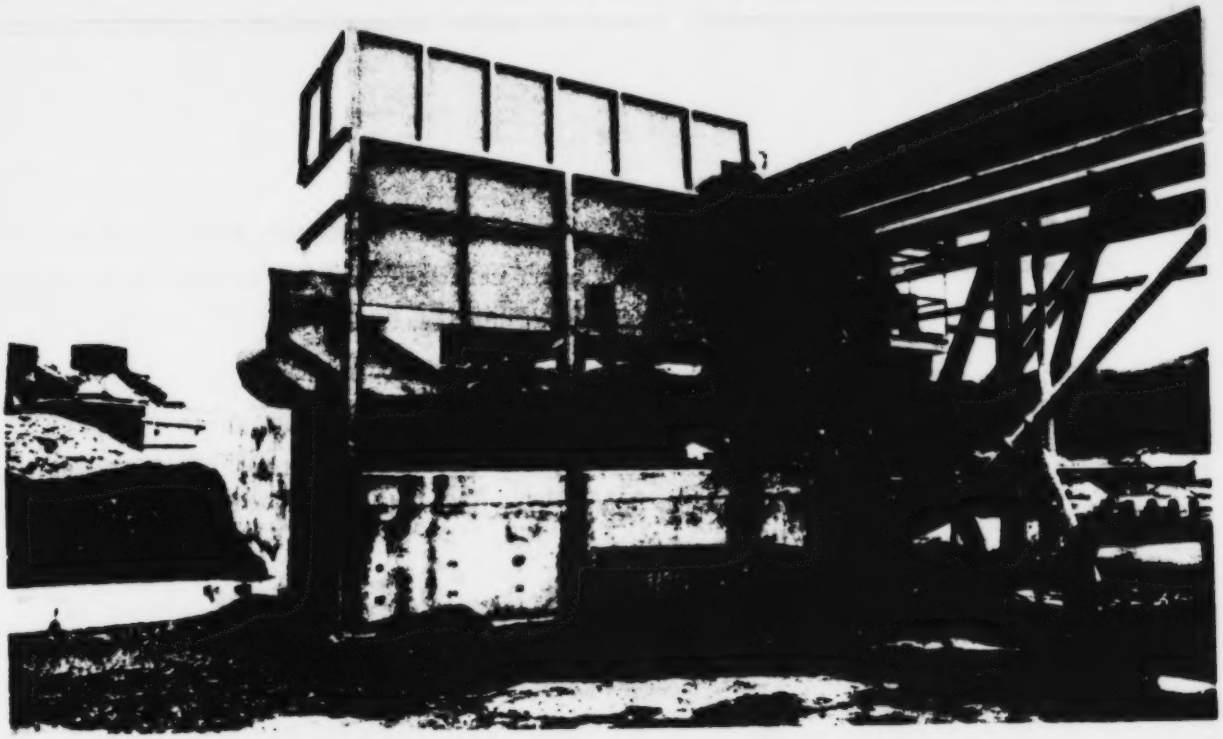


FIG. 6 METERING BIN

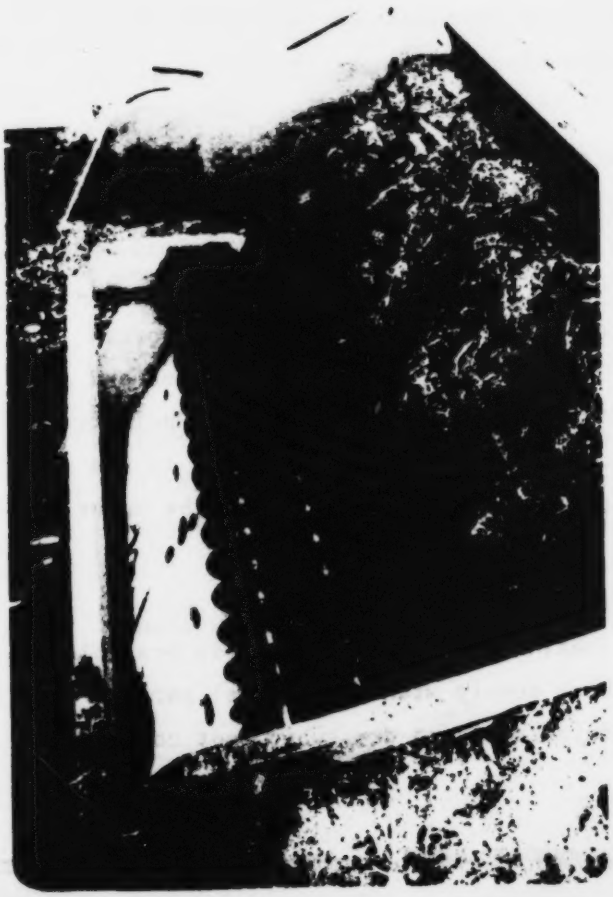


FIG. 7 DISC SCREEN AT METERING BIN DISCHARGE

where it is pyrolized and a portion of the gasified volatiles is burned. The fixed carbon consumed in this chamber supplies the heat for fuel drying and pyrolysis.

Primary air may be mixed with burner discharge gases for preheat to 315° C. (600° F.); the potential of this facility was not explored during these trials.

The percentage of volatiles burner in the primary chamber is determined by the moisture content of the wood; the wetter the wood, the greater the quantity burned. The determining factor is the primary chamber exit temperature; a minimum temperature of about 815° C. (1500° F.) is needed to ensure stable burning in the upper (secondary) chamber. In this chamber, the remaining volatiles are burned with minimal excell air. The chamber is cylindrical and double-walled with an inner refractory lining.

Combustion air, preheated by passing between the outer shell and this hot refractory, enters the chamber through a series of tangential tuyeres. As in a dust-collection cyclone, most of the air enters the upper chamber near the top, promoting a strong outer vortex of relatively cool air. This vortex spirals down inside the refractory lining to the lower end where it meets with hot volatiles entering from the primary chamber. Rapid and complete combustion takes place in the inner vortex, rising toward the central exit from the upper chamber.

By separation of gasification and combustion, it is possible to generate products which comply with the nominal particulate standard of 0.1 grains/standard dry cubic foot corrected

to 12% carbon dioxide. The fixed carbon is consumed in the primary chamber and thus supplies the heat for the remainder of the drying and pyrolysis.

At Port Alberni, the design objective was to produce 63 GJ/hour (60-MM Btu/h) from local hog fuel with a typical moisture content of 55%, wet basis.

An outlet scroll mounted on top of the secondary chamber discharges the hot gas into a horizontal, refractory-lined hot gas duct. An emergency dump valve located at the top of the scroll, combined with a block valve in the horizontal duct, permitted all hot gas from the burner to be dumped to atmosphere under upset conditions.

The hot gas duct was provided with an orifice for the purpose of comparative measurement of hot gas flow through the duct. Thermocouples were provided to monitor temperatures. Fig. 8 shows the burner at Port Alberni. The hot gas delivery duct is shown with its 90° elbow turned upward for test purposes.

Lime Kiln

Port Alberni No. 1 lime kiln is 61 m (200 ft) long by 2.13 m (7 ft) outside diameter, and is thermally rated at 52.8 tonnes/day of reburned lime. The kiln is divided into four zones.

Details are as follows:

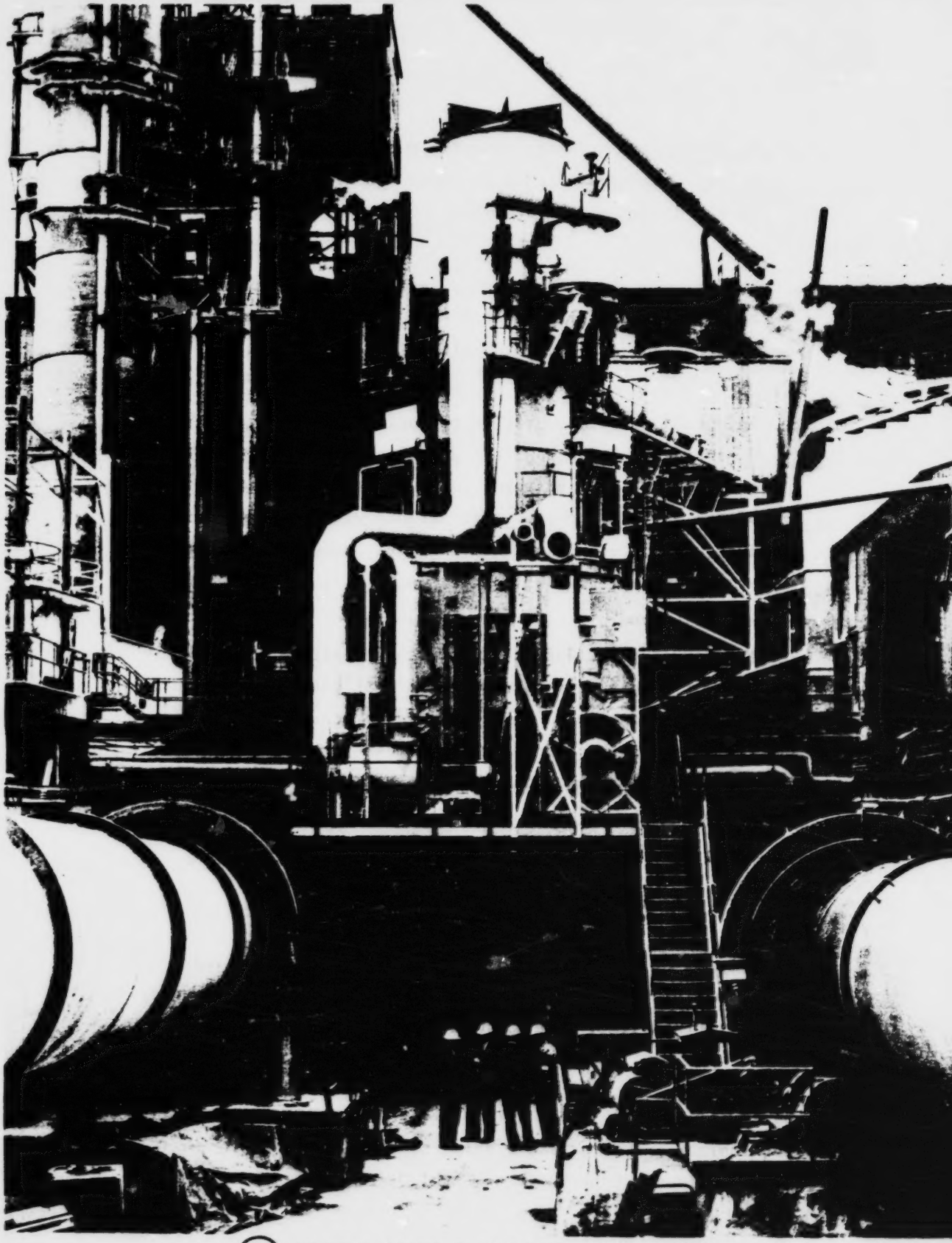


FIG 8 63 GJ/h (60-MM Btu/h) LAMB WET-CELL BURNER

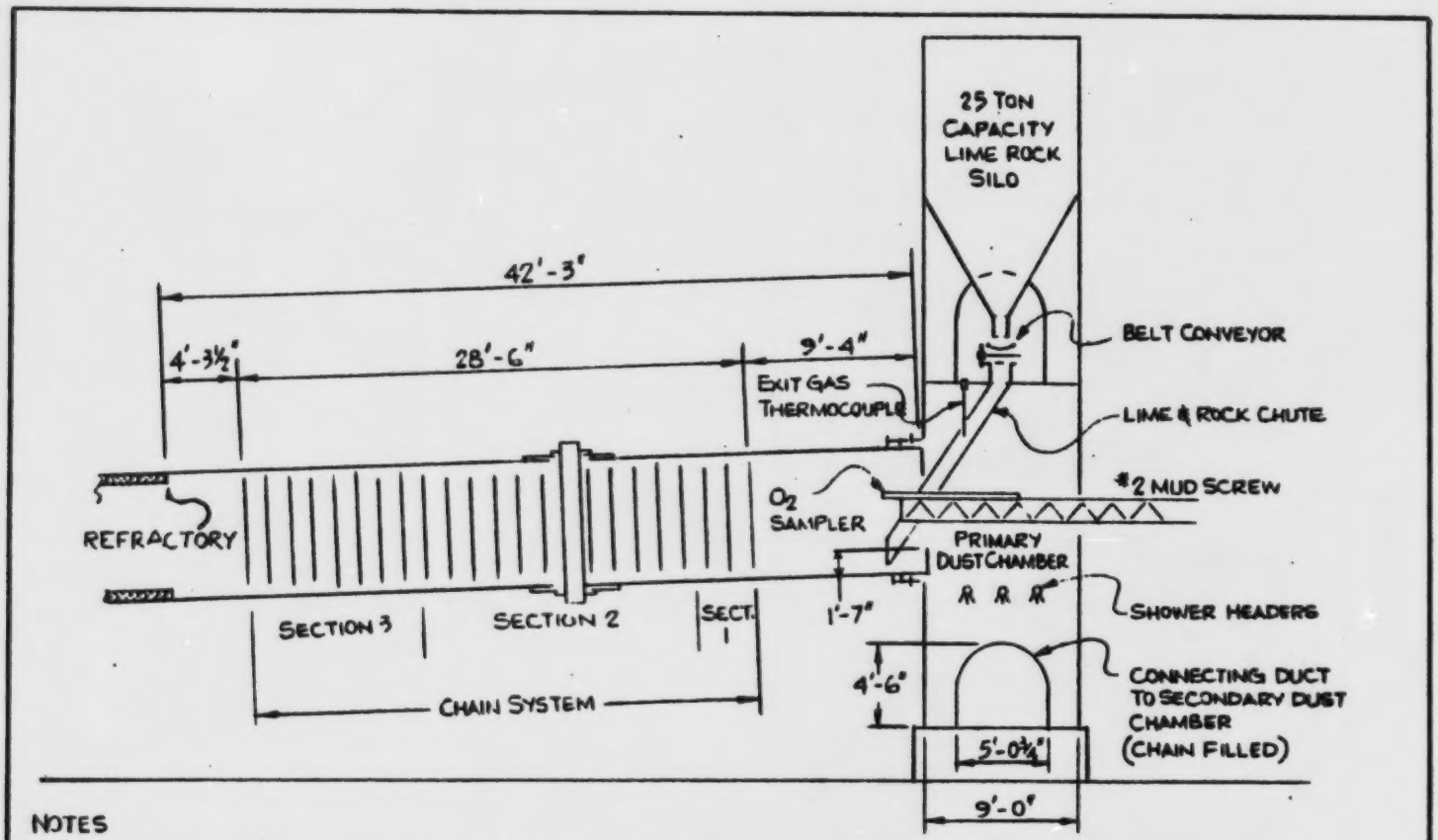
TABLE 2 PORT ALBERNI NO. 1 LIME KILN DETAILS

Section:	1	2	3	4
	Feed			Discharge
O.D. (m)	2.13	2.13	2.13	2.13
Length (m)	12.80	12.80	20.10	15.20
Refractory thickness (mm)	Nil	152.00	152.00	152.00
Refractory type	Nil	Skagit	← CB70D →	
Mean Thermal Conductivity:				
(J/sec m °C)	1.73	1.67	1.57	1.41
(Btu/h ft °F)	1.00	0.97	0.91	0.82
Maximum temperature (°C)	--	1752	← 1835 →	
Operating Range (°C)		280/770	750/1030	1010/1500

The feed end of the kiln was equipped with three sections of chain. Exit gases leaving the kiln pass downward through a primary dust chamber with showers, via a connecting duct with chain screen to a secondary dust scrubber and thence to the I/D fan. The I/D fan capacity was increased by speed change to deliver 800 m^3/min (28250 scfm) while maintaining negative pressure up to -2.5 Pa (-0.01 in W.G.) at the firing hood. The feed end is illustrated in the schematic shown in Fig. 9.

Feed slurry was dewatered by vacuum filter and fed to the kiln by two mud screws. Scrubber slurry was returned to a holding tank for recycle.

Modifications at the firing end are discussed under Trials, Section 5 (Page 43). Apart from these modifications, the operating system for the kiln, and oil/steam/combustion air controls are of standard configuration.



NOTES

1. O_2 SAMPLER – 4½" O.D. SUPPORT PIPE WITH O_2 SAMPLE PROBE PIPE INSIDE., O_2 SAMPLED 26" INSIDE FEED END FROM HOLE IN BOTTOM OF PIPE.
2. CHAIN SYSTEM – SECTION 1 – 3 ROWS., 18 CHAINS/ROW., CHAIN LENGTH 6'-3"., RING PATTERN 3" I.D., $^{25}/_{32}$" ROD.
   – SECTION 2 – 12 ROWS., 18 CHAINS/ROW., CHAIN LENGTH 6'-6"., PROOF COIL (OVAL) PATTERN., $^{5}/_{8}$" TRADE SIZE.
   – SECTION 3 – 8 ROWS., 18 CHAINS/ROW., CHAIN LENGTH 6'-3"., RING PATTERN 3" I.D., $^{25}/_{32}$" ROD.

NEW CHAIN DENSITIES (FT^2/FT^3 – SECTION 1 = 1.89., SECTION 2 = 1.53., SECTION 3 = 1.89., OVERALL = 1.72

FIG. (9) LOCATION OF PRINCIPAL FEED END ELEMENTS

LAMB - CARGATE INDUSTRIES LIMITED

System Design and Development

System Sizing

The lime kiln oil gun normally operated at 22 GJ/h to 25 GJ/h (21 to 24 x 10^6 Btu/h). The Wet-Cell Burner was rated at 63.3 GJ/h (60 x 10^6 Btu/h) using 55% M.C. hog fuel. This selection was based on early recognition that wood heat input would have to be higher, perhaps substantially, than the oil heat replaced. In addition, this provided a future option to use the same unit on one of the adjacent kilns, which were of larger capacity.

Initial testing was to be carried out at low oil substitution, so stability at low burner output was essential. The two-chamber concept permits stable operation through process control of underfire air at turndown ratios up to 5:1.

Since conversion to wood heat would obviously result in larger gas flows through the kiln for rated production, the capacity of the existing induced draft (I/D) fan was increased by approximately 40%.

Erection

Engineering commenced in March 1979. Installation of the system was extremely difficult due to the constricted job site and the necessity for steel pilings below the burner and hot gas duct. In was nevertheless completed by February, 1980, without disruption to mill production schedules.

Burner Commissioning

The burner was first fired on July 27, 1980 and throughout August, 1980. Problems were encountered in the fuel feed system due to oversize material. These were resolved by the incorporation of a rotating disc screen at the metering bin exit in September, 1980.

The burner ran again in September, 1980, and almost daily until October 30, 1980. It performed above expectations. Temperatures of 1000° C. (1830° F.) were recorded with 67% M.C. fuel (wet basis).

The hog fuel system was plagued with stoppages due to tramp metal during this period, and a magnet/conveyor belt system was incorporated at the disc screen dropout in November/ December, 1980. The system functioned well.

In January, 1981, the problem of reliable fuel again prevented tuning of the burner controls. A source was found in February which provided fuel of a more consistent quality and moisture content.

Refractory and stem seal problems on the hot gas duct block valve were resolved by April, 1981, and the vertical hot gas duct connection to the kiln hood was completed in early May, 1981.

Kiln Oil Gun

Wet-Cell gases were first admitted to the kiln on May 23, 1981. The oil flame was not stable and was difficult to sustain. The base of the flame was observed to detach itself from the fuel nozzle and to move 1 to 2 m along the kiln axis. This suggested that the Wet-Cell gases were inhibiting the mixing of combustion air with the oil vapour emerging from the nozzle.

A turbulence shield was incorporated in the annular space of the primary air inner tube next to the oil nozzle. This comprised small air "spoilers" which increased turbulence in the area and promoted better mixing of oil spray and air. Fig. 10 shows the spoilers at the exit of the primary air inner tube. This measure substantially improved flame stability and it was possible to remove the secondary air supply fan from the system. This permitted final shakedown trials to take place in November, 1981.

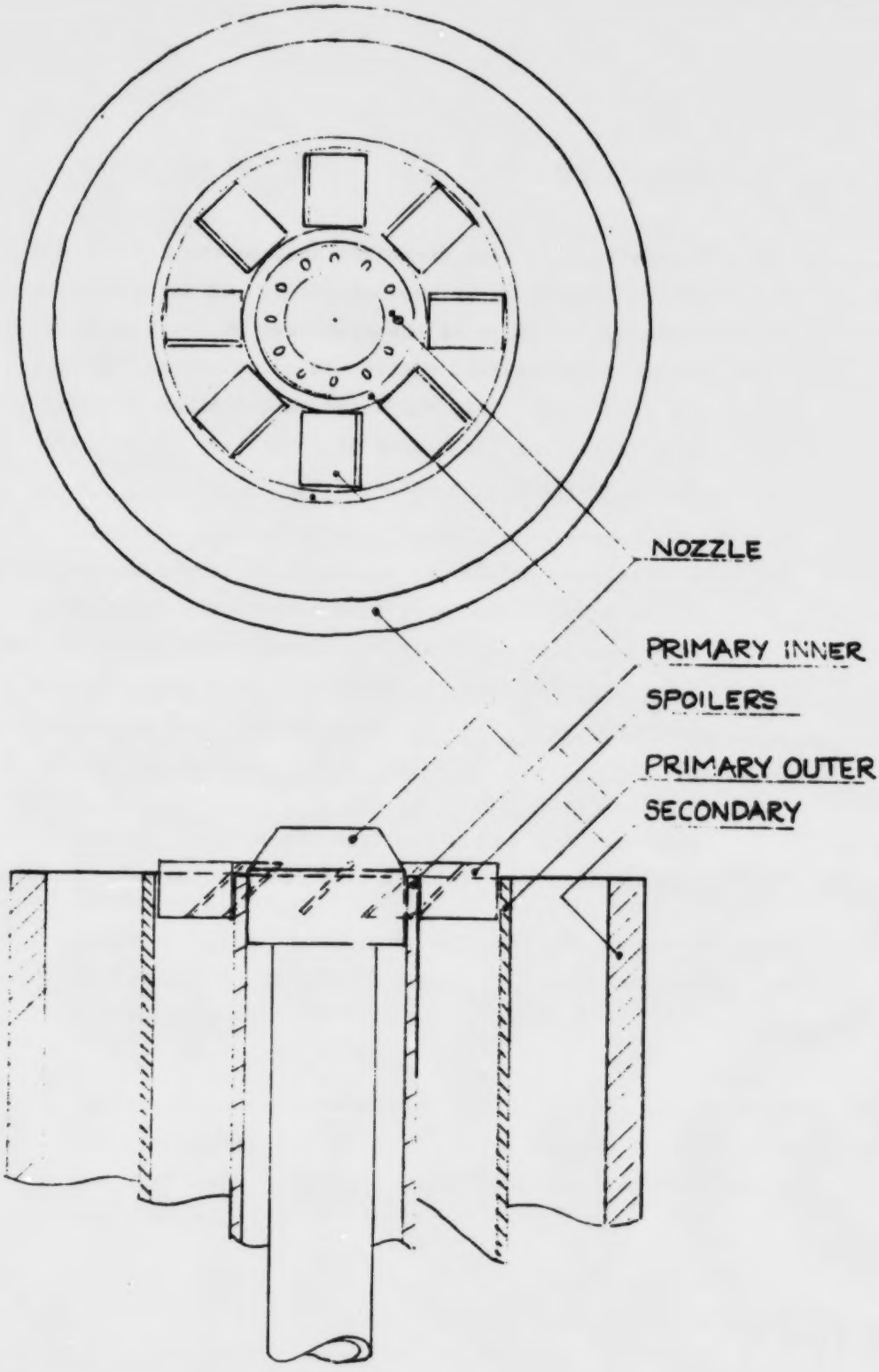


FIG. (10) TURBULENCE SHIELD IN OIL GUN

4. METHODS

The Process Flow Diagram (Fig. 11) shows the location of parameter measurements and sampling points. Additional details of the specifix measuring techniques employed were as follows:

Wet-Cell Burner

Wood Fuel Moisture Content

Percent moisture content was established by weighing before and after evaporative drying of fun-of-fuel samples. Samples were taken once per hour prior to entry to the fuel blower system. Moisture content was typically 55% (wet basis) throughout the test period.

Combustion Air Flows

Total inlet air flows (underfire, swirl and secondary) were monitored by observation of pressure drops across orifices located in the air lines. Control room water gauge pressure drop readings were calibrated prior to the start of trials using pitot traverses.

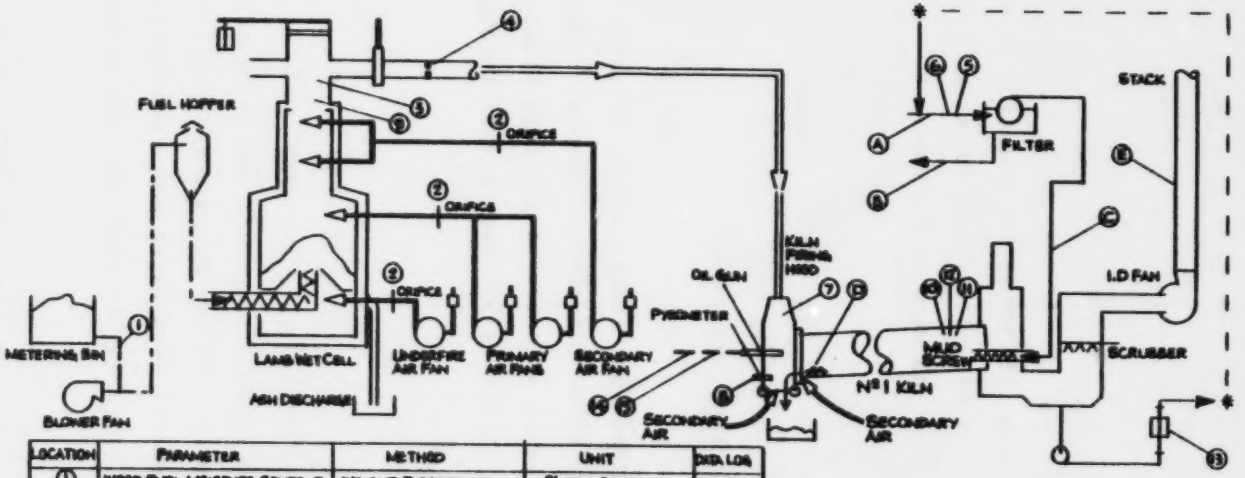
Outlet Temperature

The burner outlet temperature was measured by a calibrated thermocouple.

KILN DATA

	SECTION 1	2	3	4
	FEED			DISCHARGE
I/D	← 7'-0" (2.13m) →			
LENGTH	42.5 FT. 12.9 m	42.5 FT. 12.9 m	65 FT. 20.1 m	30 FT. 9.2 m
REFRACTORY THICKNESS	NIL	← 6" (152mm) →		
REFRACTORY TYPE	NIL	BASALT X	← CB-70-D →	
THERMAL CONDUCTIVITY J/SEC-M-°C	1.73	1.67	1.57	1.41
MAX. TEMP.	—	1752°C	← 1835°C →	
OPERATING RANGE °C	800°-1000°	1000°-1200°	1200°-1400°	1400°-1600°

SLOPE	½" PER FOOT
RESIDENCE TIME AT 1.2 RPM	122 MIN.
AT 0.7 RPM	210 MIN.
CHAIN SECTION	TOTAL LENGTH 8.68m
	OVERALL DENSITY 5.64
EXIT GAS TEMP LIMIT	650°F 345°C
MAX. EXIT GAS TEMP	700°F 371°C
EXIT GAS T/C LOCATION	1.52 m FROM GUN END
PYROMETER AIMPOINT	228mm FROM FIRING END
O_2 ANALYZER LOCATION	661mm FROM FEED END
FUEL	#6 OIL
FLAME LENGTH	7.62m
NORMAL SPEED	1.2 RPM
FIRING HOOD PRESSURE	0 TO -25 Pa



LOCATION	PARAMETER	METHOD	UNIT	DATA LOG
①	WOOD FUEL MOISTURE CONTENT	WEIGHT DIFFERENCE	% WET BASIS	NO
②	COMBUSTION AIR FLOWS	ORIFICE ΔP	"WATER GAUGE	NO
③	WET CELL OUTLET TEMP.	T/C	°C	YES
④	WET O_2	ANALYZER	% VOLUME	NO
⑤	MUD FEED	FLOW METER	US GPM	YES
⑥	MUD DENSITY TO FILTER	CHART GAUGE	kg/m³ (LB/FT³)	YES
⑦	DRAFT (FIRING HOOD PRESSURE)	BAILEY METER	Pa. ("W.G.)	YES
⑧	PRODUCT TEMP.	PYROMETER	°C (°F)	YES
⑨	WET CELL OUTPUT	INTEGRATING CONTROLLER	GJ/HR. (KW)/(BTU/HR.)	YES
⑩	EXIT GAS TEMP.	T/C	°C (°F)	YES
⑪	FLUE GAS O_2	ANALYZER	% VOLUME	YES
⑫	TOTAL RETAINED SULPHUR	ANALYZER	PPM	YES
⑬	DUST CHAMBER RETURN FLOW.	FLOW METER	US GPM	YES
⑭	OIL FLOW 1.	TOTALIZER	US GAL.	NO
	2.	FLOW METER	US GPM	YES
⑮	OIL TEMP.	GAUGE	°C (°F)	NO

LOCATION	SAMPLE POINT FOR	% SOLIDS	DENSITY	%Ca O	%Ca CO_3
Ⓐ	MUD TO FILTER	YES	YES	—	—
Ⓑ	FILTRATE	YES	YES	—	—
Ⓒ	MUD TO KILN	YES	YES	—	—
Ⓓ	LIME	—	—	YES	YES
Ⓔ	STACK	GAS COMPOSITION, PARTICULATE LOADING, GAS VELOCITY.			

FIG. (11) PROCESS FLOW DIAGRAM
SHOWING PARAMETER MEASUREMENT POINTS

Excess Oxygen

Excess Oxygen was measured continuously by a Thermox analyzer. The probe was located in the hot gas duct downstream of the orifice. This reading was used to monitor burner combustion excess air levels.

Heat Output

Btu output from the Wet-Cell was set by means of a pneumatic output controller, providing automatic adjustment of underfire, swirl and secondary air, and fuel flow.

Calculation of burner output requires measurement of the following parameters:

(a) Fuel analysis
(b) Fuel moisture content
(c) Fuel flow
(d) Ash flow and analysis
(e) Combustion air flow

Fuel and ash flows could not be continuously measured except in gross quantities difficult to ascribe to a given test period. Fuel flow, however, was derived from burner outlet temperature which was continuously monitored.

Wood chemical composition is fairly constant (5) and continuous analysis was therefore not required. Higher Heating Value (HHV) may be deduced from ultimate analysis (Table 3) by empirical formula or by calorimetry. The latter method, which gave lower values than the formula, was used for the tests.

TABLE 3 ANALYSIS OF FUEL SAMPLES

Sample:	1	2	3	4	5	Average
Higher Heating Value GJ/kg	21.43	21.24	20.71	20.74	20.70	20.97
Ash %	3.65	3.19	2.33	2.96	3.17	3.06
Carbon %	52.50	52.90	51.50	52.20	51.50	52.10
Hydrogen %	6.08	6.16	6.24	6.04	6.19	6.14
Nitrogen %	0.27	0.26	0.25	0.29	0.28	0.27
Oxygen %	38.70	38.00	39.40	38.10	38.80	38.60
Sulphur %	0.019	0.036	0.030	0.002	0.050	0.027

The enthalpies of CO_2, H_2O, O_2 and N_2 may be calculated at a given product (i.e., flame or secondary chamber) temperature by power series of the form,

$$h_i = C_1T^{n_1} + C_2T^{n_2} + C_3T^{n_3} + \text{etc.}, \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (2)$$

where h_i = enthalpy of constituent gas, and
C_1, C_2, C_3 etc., are constants, and
n_1, n_2, n_3 etc., are exponents.

From an ultimate analysis, the amount of stoichiometric combustion air per unit of dry fuel input and the proportions of the combustion products by weight may be derived. The energy released by stoichiometrically-burned hog fuel is available to heat excess air. For unit weight of a given fuel, knowledge of the operating temperature permits the excess air and hence total combustion air to be calculated.

Flow quantities (fuel and air) may then be tabulated for fixed heat output at the outlet temperature and fuel moisture content. The observed combustion air flow will then yield the actual heat output from the tabulation by simple proportion.

The Wet-Cell output refers to the enthalpy of exhaust gases above 14° C. (60° F.) with all the water in the vapour state, i.e., lower heat value (LHV).

The above calculation also yields the gas composition from which heat transport properties (viscosity, specific heat and thermal conductivity) may be derived.

Sample Calculation

Step 1

Input conditions:

Burner outlet temp	T_O	= 1160° C.
	H_2	= 6.19%
	C	= 51.5%
	O_2	= 38.8%
	Ash	= 3.06%
Ambient air temp	T_i	= 15.5° C.
HHV (dry basis)		= 20.97 MJ/kg
Fuel M.C. (wet basis)		= 0.5

Step 2

Calculate Stoichiometric Combustion Gas Proportions:

If x = gas weight/unit weight of dry fuel

$$x_{CO_2} = \frac{44}{12} \times \frac{52.5}{100} = 1.88 \quad \text{.....(3)}$$

$$\text{Required } x_{O_2} = \left[\frac{32}{12} \times \frac{51.5}{100} + \frac{8}{1} \times \frac{6.19}{100}\right] - \left[\frac{38.8}{100}\right] = 1.48 \quad \text{...(4)}$$

$$x_{N_2} = x_{O_2} \times \frac{0.768}{0.2315} = 4.91 \quad \text{.....(5)}$$

$$x_{H_2O} = \frac{6.19}{100} \times \frac{18}{2} = 0.557 \quad \text{.....(6)}$$

and x_{air} = stoichiometric air required

$$= x_{O_2} + x_{N_2} = 6.39 \quad \text{.....(7)}$$

Step 3

Calculate Pre- and Post-Combustion Gas Enthalpies by Power Series

Subscripts i = inlet or ambient conditions

o = outlet or stack condition

h_i air = 1.88 MJ/kg

$h_o CO_2$ = 1.53 MJ/kg

$h_o N_2$ = 1.60 MJ/kg

$h_o O_2$ = 1.48 MJ/kg

$h_o H_2O$ = 3.05 MJ/kg

$h_i H_2O$ = 2.44 MJ/kg (latent heat)

Fuel Low Heat Value (dry basis) = $HHV - x_{H_2O} (h_{i_{H_2O}})$ .....(8)

= 20.97 - 0.557 (2.44)

= 19.61 MJ/kg

Step 4

Calculate Flows

(a) Derive R = ratio $\frac{H_2O \text{ in fuel}}{\text{dry fuel}} = \frac{MC}{1 - MC}$ .....(9)

$= \frac{0.5}{1 - 0.5} = 1.0$

Use R as follows:

(b) Calculate heat available to raise temperature of excess air to T_o from T_i

$$Q_{AV} = LHV - R h_{iH_2O} - [x_{CO_2} h_{oCO_2} + x_{N_2} h_{oN_2} + (x_{H_2O} + 0.01 x_{air} + R) h_{oH_2O} - x_{air} h_{iair}] \quad .....(10)$$

= 3.34 MJ/kg dry fuel.

($0.01 \times_{air}$ = H_2O in atmospheric air, or humidity)

(c) Calculate heat required to raise stoichiometric air to T_o from T_i

$$Q_{st} = x_{N_2}\, h_{oN_2} + x_{O_2} h_{oO_2} + 0.01\, x_{air} h_{oH_2O} - x_{air} hi_{air} \quad ...(11)$$

$$= 8.34 \text{ MJ/kg dry fuel}$$

(d) Ratio $\frac{Q_{AV}}{Q_{st}} = \frac{3.34}{8.34} = 0.4$ = allowable excess air fraction

$$= e \quad .....(12)$$

(e) Calculate total gas flow/kg dry fuel =

$$x_{CO_2} + x_{N_2}\,(1+e) + x_{O_2} e + (x_{H_2O}+R) + 0.01\, x_{air}(1+e) \quad ....(13)$$

$$= x_{tot}$$

(f) Calculate % composition of outlet gas

eg: $\%\ CO_2 = \frac{x_{CO_2}}{x_{tot}} = 17.16\%$ .....(14)

(g) All flows may now be derived for a fixed heat output, or the heat output derived for a known flow of combustion air.

eg: Suppose Total Air Flow (from calibrated orifices) = 11300 kg/h

with $x_{air} = 6.39$ (stoichiometric air) and $e = 0.4$,

$$\text{kg dry fuel/h} = \frac{11300}{6.39\ (1 + 0.4)} \quad ..... (15)$$

$$= 1263$$

$$\text{and heat output} = 1263 \times 19.61 \quad ..... (16)$$

$$= \underline{\underline{24.77 \text{ GJ/h}}}$$

Kiln

Data Logging

Kiln parameters were recorded as follows:

TABLE 4 KILN MEASUREMENTS

Parameter	Sensing Device	Indicating Controller	Chart Recorder	Indication Only
Mud Feed	Flowmeter	Yes	Yes	
Mud Density	Ohmart	No	Yes	
Draft	Bailey Meter	Yes	Yes	
Product Temp.	Pyrometer	No	Yes	
Exit Gas Temp.	T/C	No	Yes	
Flue Gas O_2	Probe	No	Yes	
Total Retained Sulphur	Probe	No	Yes	
Dust Chamber Return Flow	Flowmeter	No	No	Yes
Oil Flow	(1) Displacement Totalizer	No	No	Yes
	(2) Flowmeter	Yes	Yes	No
Oil Temp.	Gauge	No	No	Yes
Primary Air	Orifice	Yes	No	No

Stack sampling was arranged during test periods to measure emissions and product loss. Emissions testing was carried out according to (6).

For computer simulation and steady-state observation, data logging equipment was installed to monitor on a continuous basis, product temperature, oil flow, total retained sulphur (TRS), mud flow, mud density, firing hood pressure, scrubber flow and O_2 at kiln exit.

Sampling schedules were established during each test period for lime mud, scrubber slurry, mud filtrate and lime product. These samples were analyzed for mass balance requirements, i.e., % solids, density and availability. Production level and energy consumption were derived as shown in Table 5.

TABLE 5 DERIVATION OF PRODUCTION LEVEL AND SPECIFIC ENERGY CONSUMPTION

Parameter	Derivation	Equation #
% Solids	Sample Analysis	
Solids Specific Gravity (SSG)	$\left[\frac{\%\ \text{Solids}}{100}\left(\frac{\text{Water Density}}{\text{Solid Density}} - 1\right) + 1\right]^{-1}$	17
Solid Density	Sample Analysis or Standard Value	
Solids Feed	(Mud Volume Feed to Filter)x(SSG)x(% Solids before Filter) (Filtrate Loss considered negligible)	18
Dust Loss Scrubber	(Scrubber Slurry Volume Feed)x(SSG)x(% Solids)	19
Dust Loss (Stack)	Isokinetic Sampling	
Lime Production	$\text{Solids Feed} \times \frac{1-\text{Total Dust Loss}}{1} \times \frac{1}{\left(\frac{100}{\text{Avail \%}} + 0.785\right)}$	20
Specific Energy Consumption	$\frac{\text{Total Heat Into System (LHV Basis)}}{\text{Lime Production}}$	21

Lime Availability

Availability analyses for the shakedown trials (November, 1981 to February, 1982) and initial run of the First Test Series used MacMillan Bloedel's Harmac Standard Sugar Test, in which lime is slaked in sucrose solutions at ambient temperatures. Parallel analyses by Econotech Services Ltd. using SCAN-N 25:81 (Appendix) gave results 5-7% higher. Since SCAN-N 25:81 is a standard recognized throughout the industry, the method was used in all subsequent tests.

$CaCO_3$ levels were determined by a laboratory Reburned Lime Analyzer, manufactured by Econotech Services Ltd.

Oil Flow and Heat Value

Oil flow was claibrated against controller setting by timing the flow into an empty barrel. The start and finish integrator readings were also recorded. The specific gravity of the oil was measured at 15.5° C. (60° F.) compared to water at 15.5° C. (60° F.). From this, the specific gravity was measured and API temperature/volume correction factors applied (7). HHV by this method was 42.91 MJ/kg (18450 Btu/lb). LHV was 40.52 MJ/kg (17420 Btu/lb).

Mass Balance

Negative draft at the kiln hood draws atmospheric air into the kiln via doors, observation ports and the chain conveyor trench. Combustion air for the oil gun was supplied by the primary air fan, and the secondary air fan was used to cool the rotating nose ring via a circular plenum welded to the hood. The secondary air was at a fixed damper setting and the primary air supply was modulated from the control panel.

The total air intake, and thus the mass balance, may be calculated if the following parameters are known:

(a) Wet-Cell gas flow and composition, especially O_2 concentration. (See Page 31)
(b) Kiln exit gas O_2 level (analyzer).
(c) Product mass flow. (See Page 33, Table 5)
(d) Product availability. (See Page 34)
(e) Feed % solids. (See Page 33, Table 5)
(f) Fuel flow. (See Page 32, Table 4)

The calculation is as follows:

W_o =Oil flow, kg/h
W_{st} =Stoichiometric air for oil combustion, kg/h
W_{wc} =Wet-Cell gas, kg/h
W_{H_2O}=Water evaporated from lime mud, kg/h
W_{CO_2}=CO_2 from lime mud, kg/h
W_{tot}=Total mass flow from kiln, kg/h

For typical combustion of No. 6 fuel oil:

$$W_{st} = 14.0\ W_o \quad \text{.....(22)}$$

and

$$W_{tot} = W_{wc} + W_{st} + W_{xs} + W_{H_2O} + W_{CO_2} + W_o \quad \text{.....(23)}$$

If $x = \frac{\text{actual air}}{\text{stoichiometric air}} = \frac{W_{st} + W_{xs}}{W_{st}}$ = excess air ratio,

then $$W_{tot} = W_{wc} + xW_{st} + W_{H_2O} + W_{CO_2} + W_o \quad \text{.....(24)}$$

W_{H_2O} and W_{CO_2} are known from feed % solids and lime production rate respectively.

If y = O_2 weight fraction in Wet-Cell gas
and W_{O_2} = oxygen leaving kiln, kg/h
then $$W_{O_2} = y\ (W_{wc}) + (x-1)\ W_{st} \times 0.2315 \quad \text{.....(25)}$$

Since W_{O_2}/W_{tot} is known from the kiln O_2 analyzer and y is known from Page 31, equations (24) and (25) may be solved for x, and W_{tot} may be deduced. It should be noted that for simplicity, the atmospheric humidity has been omitted.

Computer Simulation

Kiln performance is dependent on many variables, and to examine the effect of each one necessitates a greatly extended test program. A validated computer simulation of the kiln performance and thermal processes offers the possibility of maximizing the information derived from a finite data base.

The data acquisition and simulation techniques developed by the Boeing Company were contracted for this study. These techniques were developed in the 1960's and have been applied as follows:

	Type of Plant	Number of Kilns	Year Completed
KILN CONTROL SYSTEMS			
Crown Zellerbach Wauna, Ore.	Pulp	1	1979
Louisiana-Pacific Samoa, Cal.	Pulp	1	1980
Northwest Alloys Addy, Wash.	Magnesia	2	1981
Crown Simpson Eureka, Cal.	Pulp	1	1982
KILN CONSULTING STUDIES			
American Can, Halsey, Ore.	Pulp	1	1979
Kaiser Refractories Natividad, Cal.	Dolomite	3	1981
Lamb-Cargate	Pulp	1	1977
MacMillan Bloedel Port Alberni, B.C.	Pulp	2	1978

KILN CONSULTING STUDIES	Type of Plant	Number of Kilns	Year Completed
Owens-Illinois, Valdosta, Ga.	Pulp	1	1979
St. Regis, Tacoma, Wash.	Pulp	2	1978
Stauffer Chemical Silver Bow, Mont.	Phosphate	2	1982
Weyerhaeuser, 4 Locations	Pulp	4	1975-81

The kiln simulation described below was developed by Boeing for oil- and gas-fired kiln analysis. It consists of mathematical models of the heat and material transfers and reaction kinetics which take place within the kiln. The models are based on partial differential equations which describe the state of the kiln at any given point and instant of time. In the simulation, partial derivatives of material temperatures and concentrations are computed with respect to length along the kiln longitudinal axis. These partial derivatives are a function of the current state at the point of evaluation and the time rate of change of state at that point. The time rates of change are derived by numerically differentiating at the point of interest. Length partial derivatives are then computed and integrated. The simulation is functionally described in Fig. 12. Solids and gas input conditions are computed from kiln operating conditions. The counter-flow nature of the kiln requires an initial estimate of all gas concentrations and temperatures and kiln shell temperatures which affect the solids material and energy balance equations. The profiles serve as a starting point for the simulation, and errors in the profile estimated are damped out through an iterative process. The simulation begins at the solids feed end and computes wall temperatures, solids reaction rates, and solids energy balances at the feed end. From this information and estimated profiles, the partial derivatives of the state variables with respect to length are computed. The

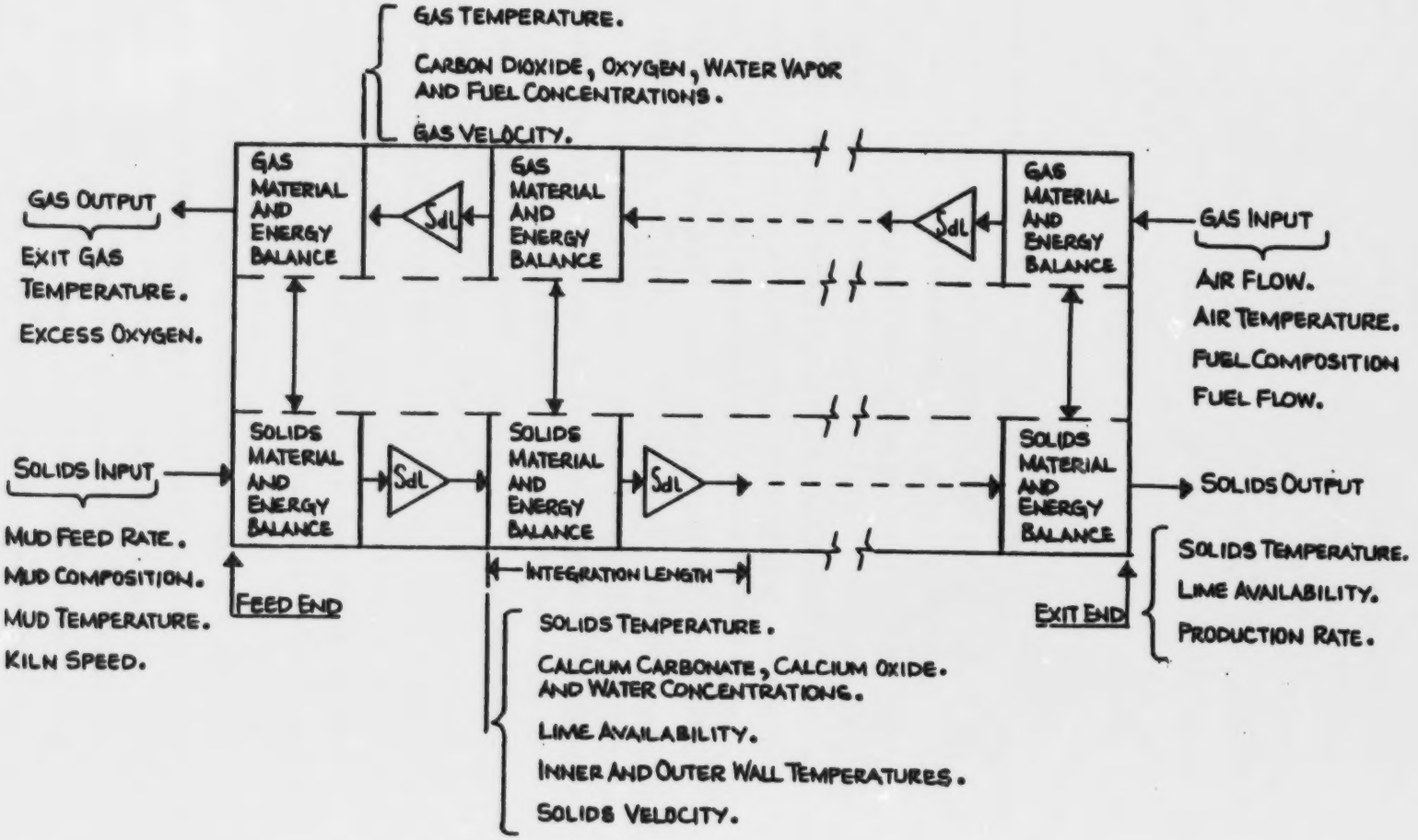


FIG. (12) LIME KILN SIMULATION FUNCTIONAL DESCRIPTION.

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derivatives are then integrated over a 2.5-foot interval to obtain the solids state 2.5-feet further down the kiln. This process is repeated until the exit end is reached. Starting at the exit end, the gas partial derivatives with respect to length are then computed and integrated until the feed end is reached. Both the solids and gas profiles are saved at five foot increments for use during the next pass. Between passes, time is incremented. During the next pass, the computed value at each point is compared with the last value at that point and the difference divided by the time interval to get the partial derivatives with respect to time. If steady-state operation is desired, the time derivatives are forced-set to zero and the simulation is operated until the profiles converge, indicating the steady-state value.

During operation, gas and solids temperature and concentration profiles are printed out at selected intervals. Solids concentrations are normalized with respect to the equivalent calcium oxide (CaO) content of the solids and gas concentrations are normalized with respect to the nitrogen (N_2) content of the gas. Additionally, availability of calcium oxide is computed and printed out.

The simulation allows the kiln to be subdivided into up to three or four characteristic zones. For a typical kiln, the first zone defines the chain section or drying zone, the next zone is the calcining zone, and the last zone is the burning zone. These terms are distinguished by distinct heat transfer parameter values and inside diameters due to varying refractory thicknesses. These parameter values are specified as input conditions. Composite gas and solids heat capacities and gas emissivities are computed by the simulation as a function of compositions and temperatures.

Heat transfer areas are computed, as well as gas velocity profiles, throughout the kiln.

Typical simulation inputs are listed in Table 6. Currently, operating conditions are controlled manually. However, the simulation is designed to allow automatic control loops to be added to evaluate control system design. Solids and gas feed conditions are provided by separate subroutines which can be modified to control feed conditions as a function of kiln measurements and desired operating conditions.

Data acquired from the kiln are used to validate the simulation. Simulation parameters are adjusted to match actual kiln operation as closely as possible. Two approaches are used during the validation process. First, simulated profiles are compared to actual kiln data for steady-state operating conditions. Then simulation sensitivities to parameter variations such as fuel and mud feed rates are compared with data obtained from the kiln. Parameter values are refined to obtain agreement with kiln test data. If total air flow data are not available from the kiln, the simulation airflow is obtained by matching excess oxygen values.

Although the lime kiln simulation was developed primarily for use in control system design and evaluation, the simulation may be used to evaluate operating procedures for existing installations to improve efficiency, or may be used in kiln design to evaluate the effects of varying design parameters, such as kiln dimensions, refractory selection, air flow requirements, and burner selection. For this trial, the simulation provided detailed information on the process behavior throughout the kiln length which was not obtained in actual kiln operation.

TABLE 6 KILN SIMULATION INPUT QUANTITIES

KILN PHYSICAL PARAMETERS:

- Solids Emissivity (each section)
- Inner Wall Emissivity (each section
- Outer Wall Emissivity (each section)
- Gas -- Solids Convective Heat Transfer Coefficient (each section)
- Gas -- Wall Convective Heat Transfer Coefficient (each section)
- Wall -- Solids Convective Heat Transfer Coefficient (each section)
- Inner -- Outer Wall Convective Heat Transfer Coefficient (each section)
- Wall -- Ambient Convective Heat Transfer Coefficient (each section)

- Inside Diameter (each section)
- Outside Diameter (each section)
- Kiln Slope
- Solids Residence Time Factor
- Fuel Combustion Rate

KILN OPERATING CONDITIONS:

- Solids Feed Temperature
- Ambient Air Temperature
- Primary/Secondary Air Temperature
- Solids Feed Rate
- Fuel Flow
- Air Flow
- Feed Moisture Content
- Maximum Calcium Oxide Availability
- Kiln Rotational Speed

SIMULATION RUN PARAMETERS:

- Integration Length Interval
- Integration Time Interval
- Run Time
- Print Time Interval
- Special Diagnostic Print
- Request and Print Time
- Run Mode -- Steady-State or Transient

5. TRIALS

The trials can be divided into three chronological groupings as follows:

November, 1981 - February, 1982
Equipment shakedown and modification

March, 1982
Performance testing for continuous operation to establish equipment reliability, oil baseline operation and minimum substitution level for the subsequent program.

April, 1982 - June, 1982
Detailed performance evaluation including data acquisition for computer model simulation; cause and effect testing; model validation and performance limits.

Equipment Shakedown and Modifications

Trials

Shakedown trials were carried out in November and December, 1981. The November tests showed approximately 66% oil replacement at a nominal feed rate of 40 tonnes (44 tons) lime per day. Fuel moisture content was in the range 50-60% wet basis, and wood heat quantity was estimated at 21.2 to 26.4 GJ/h (20-25-MM Btu/h). Lime availability by the Harmac sugar test varied between 75-85%, equivalent to 80-90% by SCAN-N 25:81. In December, the feed rate was increased to the normal equivalent of 45 tonnes (50 tons) lime per day. Due to miscellaneous setbacks including shortages of hog fuel and kiln drive problems, target production range

and high substitution were not achieved until December 9 and 10. Difficulties were met in supplying mud at demand rate due to feed screw motor overload. This was partially overcome by the introduction of water to the feed screw, effectively decreasing feed % solids content and lowering exit gas temperatures. The estimated Wet-Cell output during the period 0900 - 1400 hours on December 10 was greater than 31.6 GJ/h (30-MM Btu/h), concurrent with satisfactory lime production and higher than normal I.D. fan temperatures.

The shakedown trails indicated the necessity for the following modifications which were implemented prior to the second set of trials in March, 1982.

Modifications to Kiln Nose Ring

The increased mass flow from the Wet-Cell caused heating of the kiln nose ring area to outside surface temperatures in excess of 343° C. (650° F.). In extreme circumstances, positive draft in the hood produced intermittent blowout of gas through the clearance annulus where the kiln enters the kiln hood. Accordingly, air from the secondary fan was ducted to a circular plenum seal as shown in Fig. 13. The air produced a cooling effect and eliminated blowout.

Modifications to Kiln Hood

The duct carrying the Wet-Cell gases entered the hood from the top as illustrated in Fig. 14. It was found necessary to instal refractory as shown in Fig. 15. This conferred several advantages as follows:

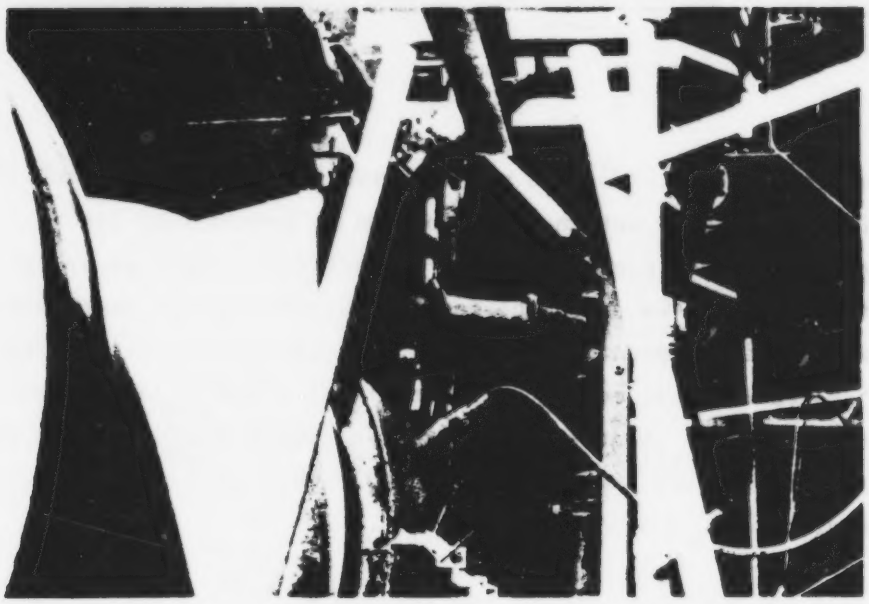
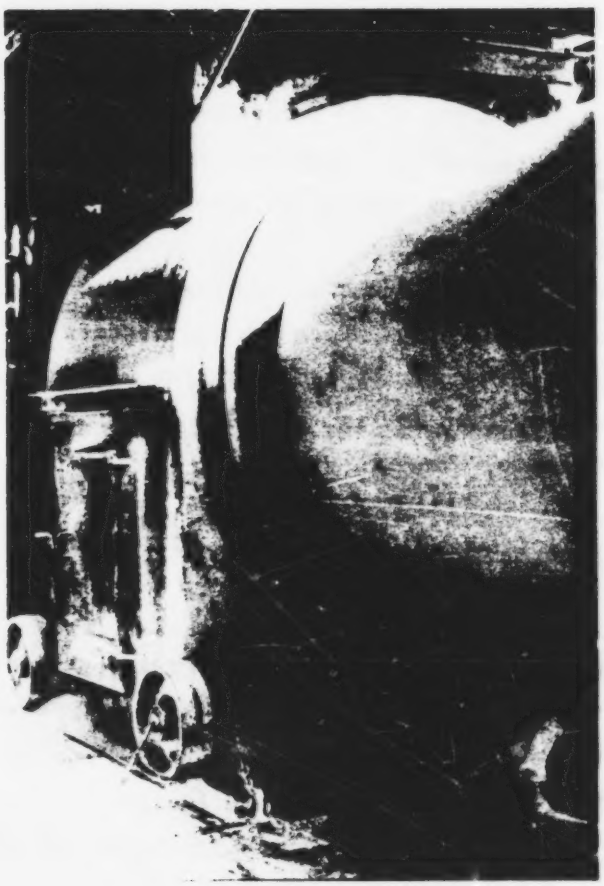


FIG. 13 NOSE RING SEAL



(a) The refractory arch over the conveyor chute (upper photograph) prevented impingement of Wet-Cell gases on the chain conveyor system, reducing product dust entrainment and temperature levels in the chain system.

(b) Metal temperatures on the oil gun were reduced. These may otherwise have led to distortion of the air pipes and oil gun support tube, preventing oil gun removal.

(c) Wet-Cell gases were directed towards the kiln axis and turbulence reduced in the space around the oil flame. The brickwork may have imparted a peripheral spin to the gases and improved flame stability by providing an axial vortex in which oil combustion could proceed.

(d) The refractory protected kiln hood metal and fittings and reduced surface metal temperatures.

The above measures contributed substantially to flame stability and to the system's capability for sustained performance. Comments with regard to future design are presented in Section 9.

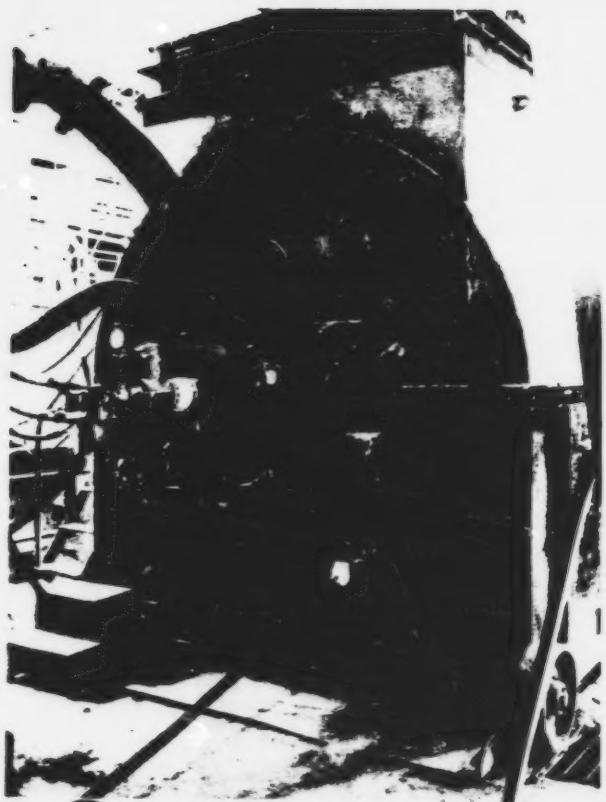


FIG. 14 KILN HOOD SHOWING TOP GAS ENTRY

FIG. 15 REFRACTORY MODIFICATIONS IN KILN HOOD

Performance Testing

First Test Series

Normal Operation -- 100% Oil-Firing

The kiln functioned with operator-determined control-settings from 0900 March 11 to 0500 March 13; a duration of 44 hours. Mud feed rate was set at 5.90 m^3/h (26 USGPM) or 76-78 tonnes (84-86 tons) solids/day.

Data are presented as follows:

Data profiles vs. test duration -- Fig. 16.
Production data summary -- Table 7.

Baseline Operation -- 100% Oil-Firing

The kiln was operated with controls set to minimize O_2 levels at the same feed rate of 5.90 m^3/h (26 USGPM) or 76-78 tonnes (84-86 tons) solids/day. The test ran from 0900 March 16 to 2000 March 17, a duration of 40 hours.

Data are presented as follows:

Data profiles vs. test duration -- Fig. 17.
Production data summary -- Table 7.

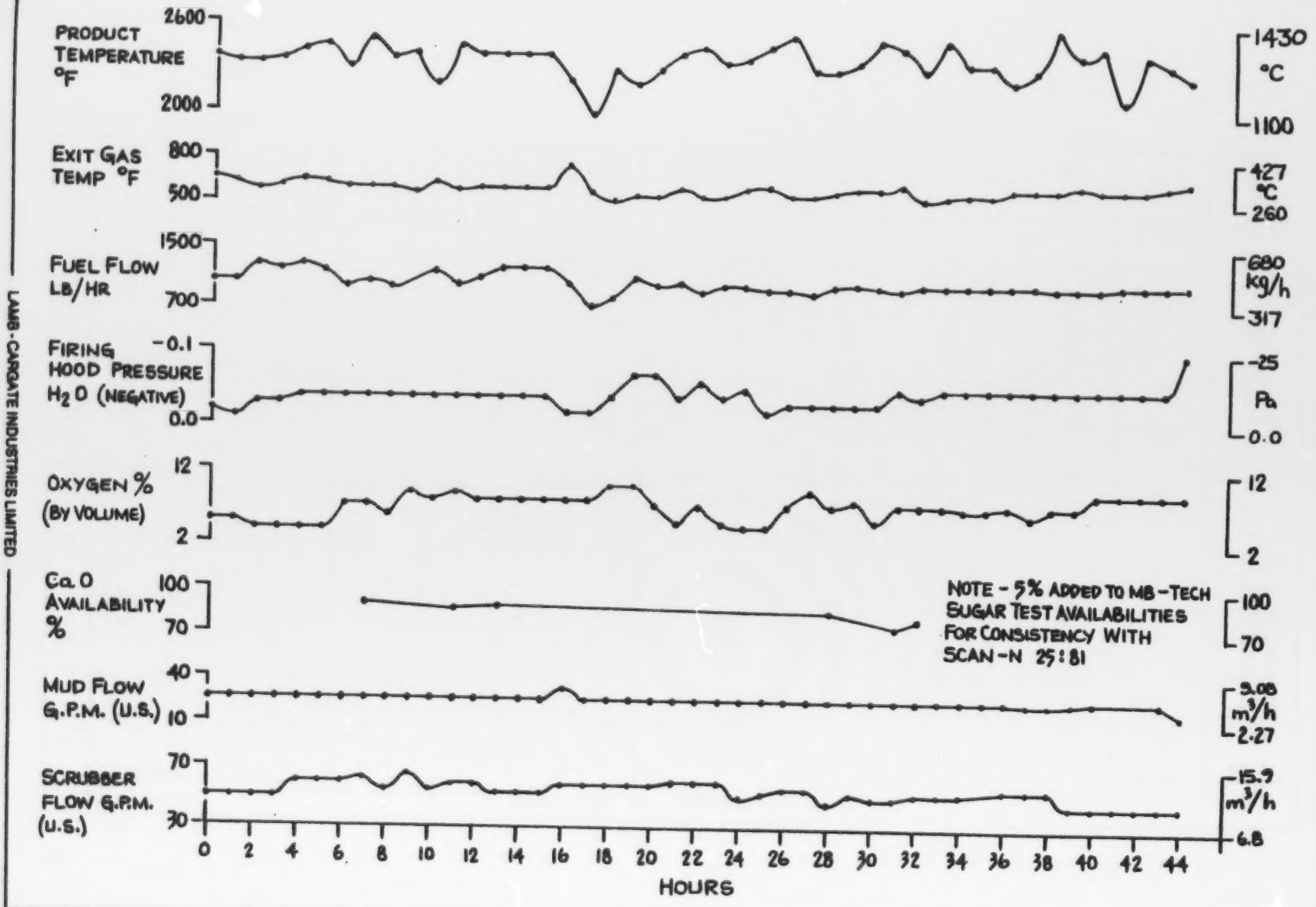


FIG. (16) FIRST TEST SERIES DATA PROFILES VS. TEST DURATION FOR OPERATOR CONTROLLED OPERATION

TABLE 7

FIRST TEST SERIES. PRODUCTION DATA SUMMARY

Hour	Feed				Scrubber Dust					Temps. °C.		% O_2 By Vol.	Availability % CaO	Oil Fuel kg/h	Production CaO tPD	Test	Wet-Cell Output GJ/h	Specific Energy GJ/t
	m^3/h	% Solid	Sp.Gr.	tPD	m^3/h	% Solid	Sp.Gr.	tPD	% Loss	Product	Exit Gas							
7	5.90	41.5	1.35	79.6	14.1	1.49	1.009	5.08	6.38	1371	315	7.5	*90.3	481.7	39.4	100% Oil Operator Control	Nil	11.87
11	"	40.3	1.34	76.6	13.2	2.16	1.013	6.92	9.04	1343	310	9.0	*88.1	462.2	36.3			12.37
13	"	40.3	1.34	76.6	12.0	2.23	1.010	6.53	8.53	1315	315	8.0	*85.8	558.8	35.9			15.12
28	5.81	40.2	1.34	75.1	10.9	4.31	1.010	11.54	15.43	1260	304	7.5	*80.4	462.6	31.3			14.36
31	5.90	40.1	1.34	76.3	11.3	3.11	1.020	8.64	11.33	1343	315	5.5	*73.7	417.7	31.5			12.85
32	5.86	40.5	1.34	76.5	11.8	2.05	1.070	5.89	7.69	1260	282	7.5	*79.8	456.6	34.6			12.82
5	5.9	40.3	1.34	76.6	12.0	2.01	1.010	5.88	7.68	1315	232	4.5	92.9	410.9	38.0	100% Oil Baseline		10.51
7	5.81	40.2	1.34	74.9	12.71	2.01	1.010	6.21	8.27	"	260	3.0	91.4	"	36.7			10.85
12	"	40.1	1.34	"	11.8	1.09	1.007	3.10	4.15	"	204	5.5	91.4	462.2	38.2			11.72
22	"	40.1	1.34	"	12.5	1.09	1.007	3.29	4.39	1343	288	4.0	94.1	"	38.7			11.60
24	"	40.1	1.34	"	"	1.09	1.007	"	4.39	1248	343	3.5	94.1	"	"			"
26	"	40.1	1.34	"	"	1.09	1.007	"	4.39	1204	"	4.5	93.8	"	"			"
28	"	40.1	1.34	62.9	"	1.09	1.007	"	4.39	1371	332	4.0	94.1	436.6	"			10.95
32	5.27	37.9	1.31		11.8	1.74		5.00	7.90	1315	338	6.0	83.4	443.2	29.3			14.68
1	5.81									1048	"		93.5	231.3				
4	"									1082	310			"				
5	5.72	40.8	1.34	75.4	8.85	6.24	1.040	13.8	18.30	1193	315	7.5	84.6	237.7	31.4	50% Oil Substitution	24.5	25.65
9	5.63	40.0	1.34	72.4	10.9	7.12	1.040	19.5	27.00	1093	"	5.0	93.4	"	28.4		26.6	30.05
11	5.90									1126	"		94.5	244.0				
13	5.72	39.8	1.33	73.0	9.08	3.65	1.020	8.15	11.20	1071	321	8.5	85.9	"	33.3		26.2	25.5
17	5.99									1082	315		93.6	385.5				
23	5.90									1287	338		92.3					
24	"									871	354		80.7	231.3				
29	"	39.6	1.33	74.8	8.63	10.30	1.070	22.8	30.50	898	327	7.5	56.6	"	20.4		Out	10.32
31	6.09	38.3	1.31	73.8	9.08	10.70	1.070	25.9	33.90	954	354	8.0	72.1	"	22.4		Out	9.40
33	5.90									926	338		85.8	"				
36	5.72	39.3	1.33	71.7	(")	2.21	1.010	4.88	6.80	1037	"	8.5	65.5	"	28.9		23.8	27.0
37	5.90									1027	343		84.8	"				
39	"									982	321		83.3	"				
41	"	(40.4)	1.34	76.8	9.99	(3.00)	1.020	7.33	9.50	"	349	7.0	94.3	"	37.6		26.0	22.5
43	5.81	40.4	1.34	75.9	9.08	4.25	1.030	9.54	12.50	987	315	9.0	69.4	"	29.8		22.7	25.3
45	5.90	(40.4)	1.34	76.8	10.4	(4.30)	1.030	11.1	14.40	982	321	5.5	74.3	"	30.8		21.9	24.2
47	5.72	40.3	1.34	74.1	7.36	4.38	1.030	7.81	10.60	1004	327	7.0	72.5	"	30.6		26.4	27.5
49	5.90	(40.4)	1.34	76.8	6.81	(4.30)	1.030	7.18	9.40	1027	"		93.5	"	37.5		23.3	20.9

( ) Assumed Value

* 5% added for equivalence to SCAN-N 25:81

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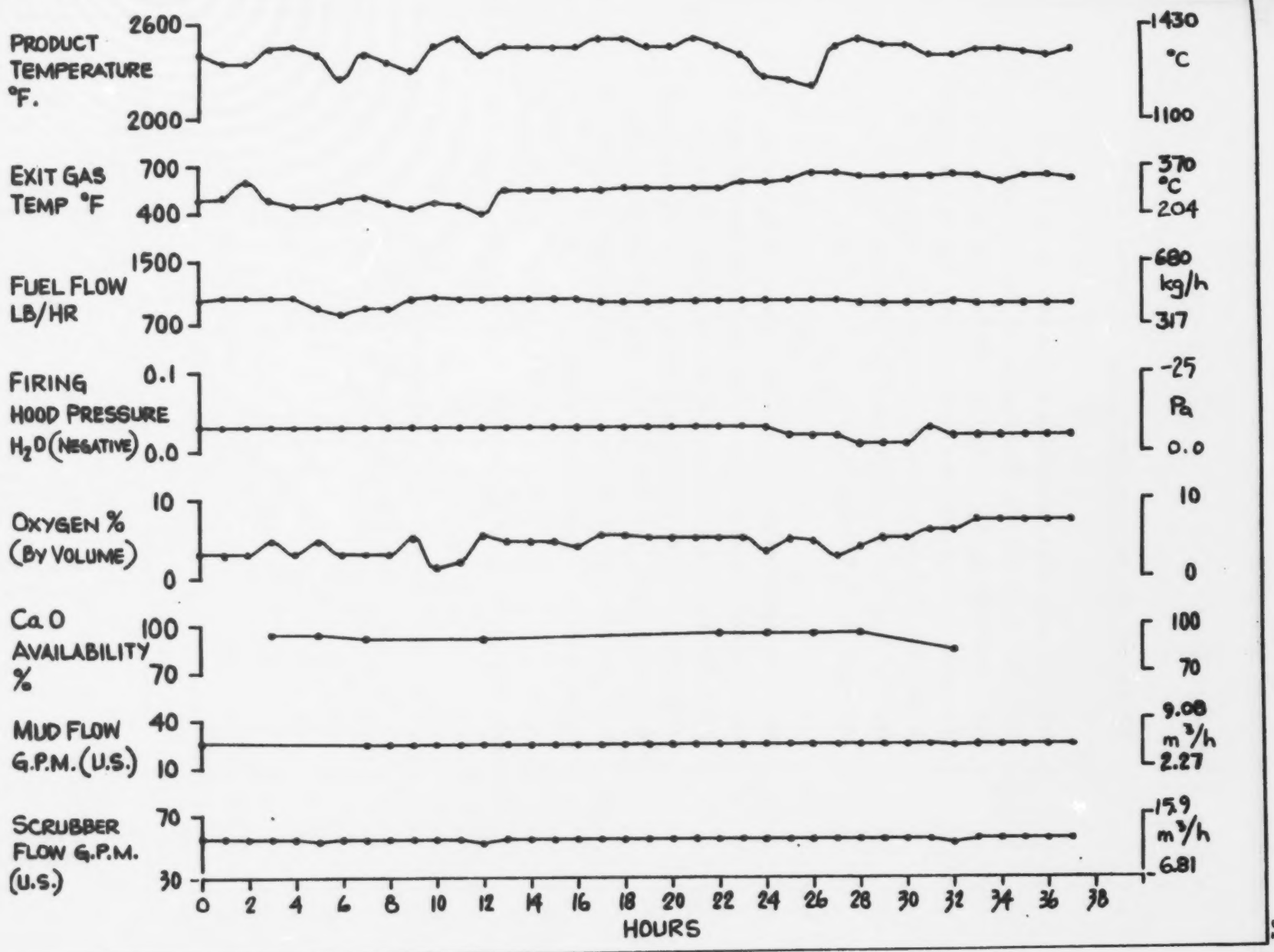

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FIG. (17) FIRST TEST SERIES DATA PROFILES VS. TEST DURATION FOR BASELINE OPERATION

50% Nominal Oil Substitution

Wet-Cell gases were introduced to the kiln at 1215 hours on March 19 and the oil flow reduced to 75% of baseline, i.e., from average 452 to 326 kg/h (998 to 720 lb/h). This situation was held for 22 hours, then oil was reduced to 231 kg/h (510 lb/h) with a corresponding increase of Wet-Cell input to 24.6 GJ/h (23.2 x 10^6 Btu/h). This corresponded to 47.9% substitution.

The trial continued for a further 53 hours and was terminated by the mill's requirement to feed limestone to the kiln. The trial was interrupted by a Burner feed problem between hours 17 and 24, during which oil flow was increased to baseline magnitude, and between hours 28 and 29, during which the abort valve was open. The effects on availability may be observed in the data profile.

Data are presented as follows:

Data profiles vs. test duration -- Fig. 18.
Production data summary -- Table 7.

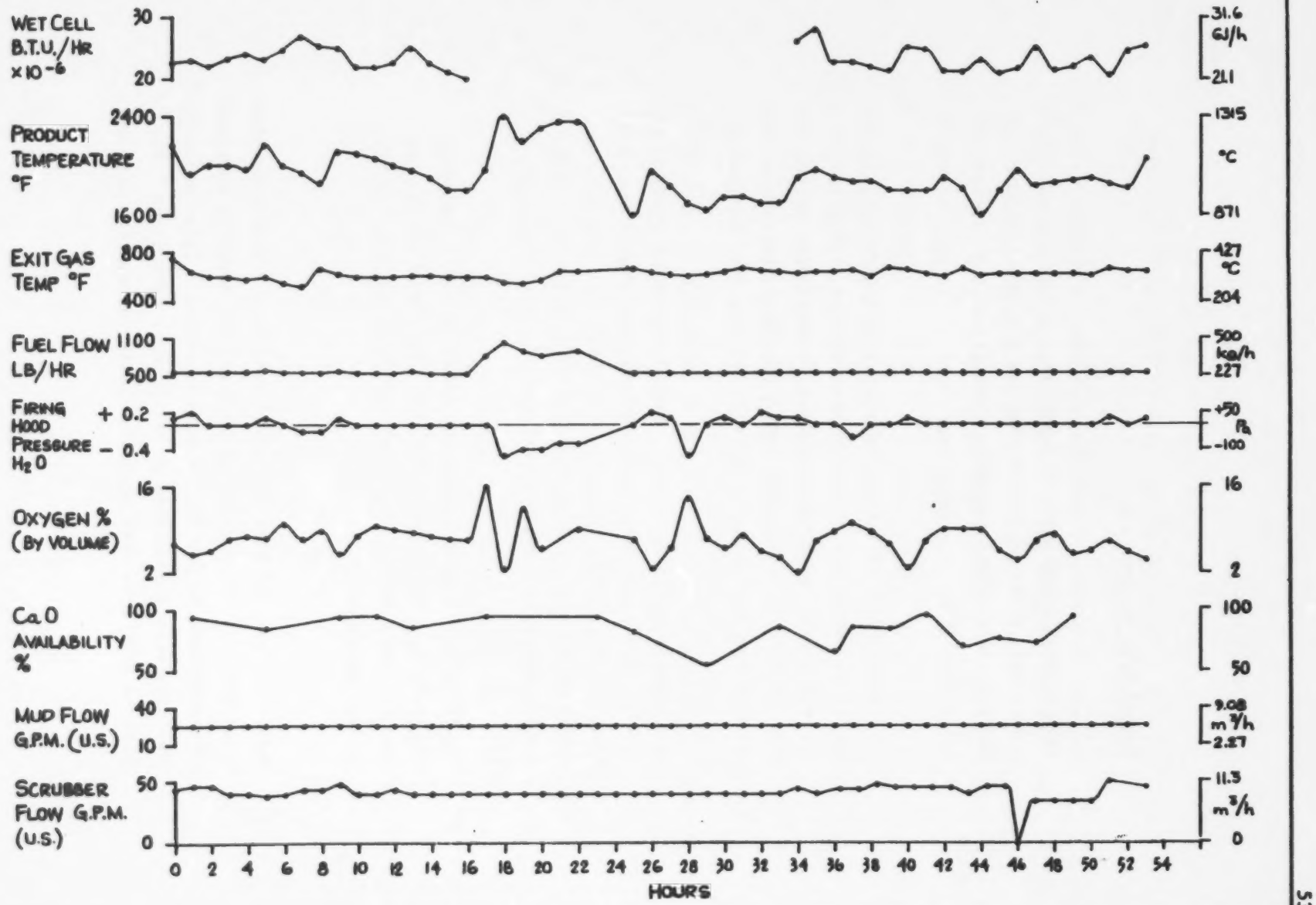

FIG. (18) FIRST TEST SERIES -- DATA PROFILES VS. TEST DURATION FOR 50% OIL SUBSTITUTION

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Second Test Series

Having established baseline operating conditions and a model simulation, a Second Test Series was performed in May, 1982. The purpose of the tests was to acquire data at increasing oil substitution.

At a given level of substitution, Wet-Cell output could be varied to assess the effect on availability, temperatures, O_2 level, etc., with the objective of developing an indication of minimum replacement ratio (Wet-Cell heat – baseline oil heat replaced) consistent with satisfactory production.

Data were recorded on a Varian Data Acquisition System, permitting continuous observation of product temperature, fuel flow, kiln O_2 level, mud flow, scrubber flow, Wet-Cell gas temperature, etc., and providing a firm statistical basis for average values selected as input to the simulation.

Ten tests were run at increasing substitution, comprising two tests near baseline, three near 50% substitution, two near 60% substitution, two near 75% substitution, and one at 100% substitution.

Tests were run for five to eight hours once stable conditions were observed. Concurrent shell scans and sampling routines were continued as for the first test series. Cause and effect test data were accumulated in strip chart recordings. The observed dust loss data showed no clear correlation with mass

velocity. This point is discussed in Section 6.

Since stack dust loss data were not concurrently observed, nominal quantities were applied to the mass balance assuming that losses increased with increasing substitution. These quantities were varied linearly from 38 kg/h (83 lb/h) at zero substitution to 128 kg/h (284 lb/h) at 100% substitution.

Data are presented as follows:

Data profile vs. test duration -- Figure 19.
Production data summary -- Table 8.

The data profile shown is for Test 5. It illustrates the smooth characteristics corresponding to steady-state conditions and was typical of the other tests in the series.

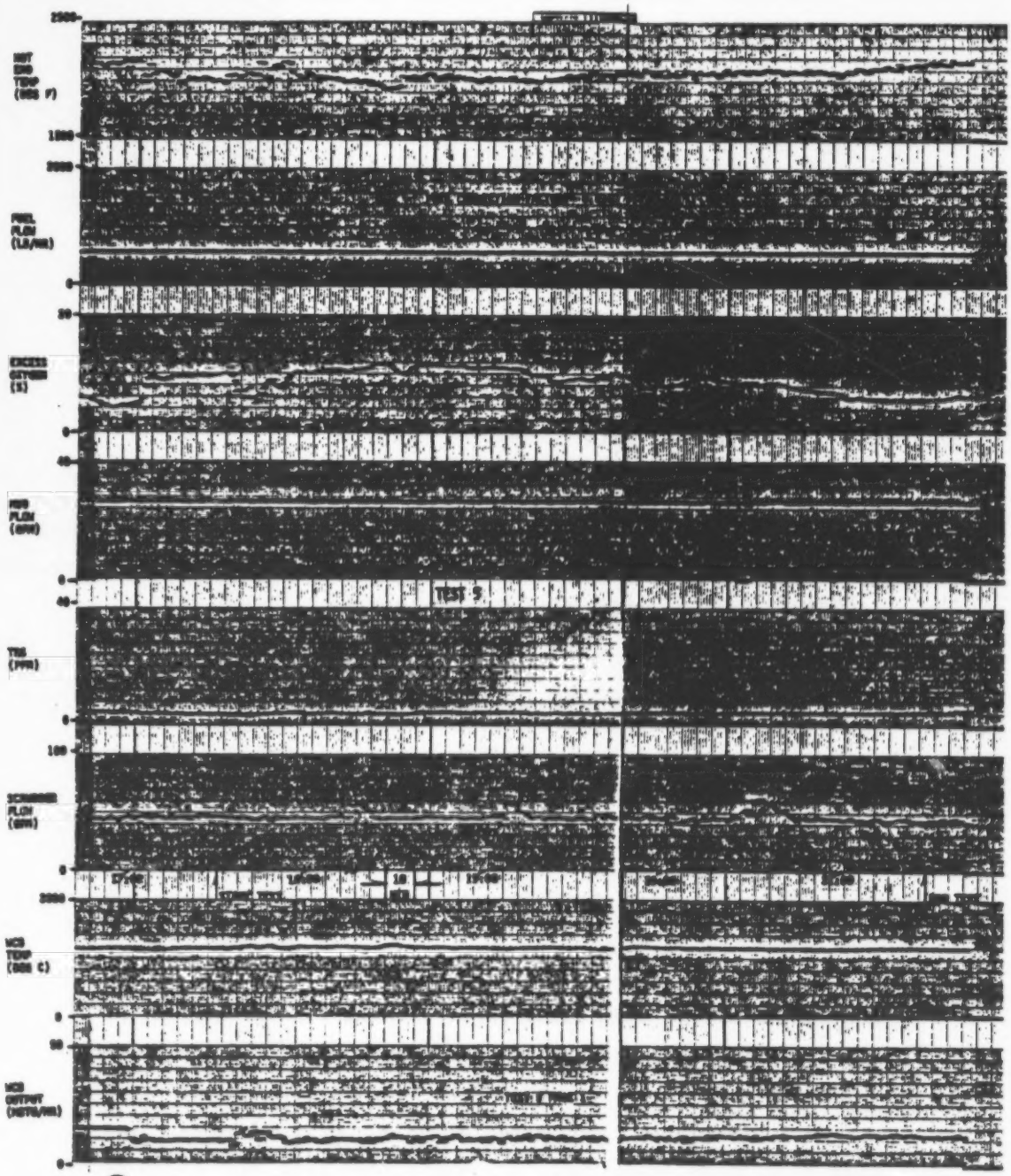

FIG (19) TYPICAL DATA STRIP-OUT -- TEST 5 -- SECOND TEST SERIES

TABLE 8 SECOND TEST SERIES -- DATA SUMMARY

Test	Date	Feed to Filter			Dust Loss (Scrubber)				Net Feed	Temperature		$\%$ O_2 By Vol.	Avail. %	Oil Fuel kg/h	CaO tPD	Specific Energy (LHV) GJ/t	Screw Water Added kg/h	Wet-Cell		% Oil Sub
										Product	Kiln Gas Out							Gas Out °C	Output GJ/h	
		m^3/h	% Solids	tPD	m^3/h	% Solids	tPD	% Loss		°C	°C									
Baseline	--	5.90	40.1	76.3	--	--	6.5	8.6	69.8			6.0	92.0	453	37.3	11.7	0			0
T-1	5/05	5.99	38.9	74.4	18.2	0.50	2.2	4.0*	71.4	1399	215	2.0	94.2	374	38.6	9.42	0			0
T-2	5/06	5.99	31.4	56.4	19.1	1.17	5.4	11.0*	50.2	1415	315	10.2	94.7	374	27.3	13.3	0			0
T-3	5/07	5.81	40.1	75.1	16.6	4.40	18.0	26.6*	55.1	982	354	9.4	49.2	235	19.5	32.4	435	1120	16.8	48
T-4	5/09	5.99	40.3	77.9	10.2	2.21	5.5	9.7*	70.3	1171	332	6.0	93.0	235	37.8	21.9	2021	1200	25.0	48
T-5	5/09	5.99	40.3	77.9	10.2	2.21	5.5	9.7*	70.3	1149	360	6.5	94.1	235	38.1	18.9	2021	1200	20.5	48
T-6	5/10	5.81	40.6	76.2	10.0	3.50	8.6	14.6*	65.1	1132	343	8.4	93.4	125	35.1	26.4	2748	1240	33.5	72
T-7	5/15	5.90	32.5	58.1	15.9	0.38	1.4	5.9*	54.7	1088	321	6.2	92.6	186	29.3	24.5	1476	1160	22.4	59
T-8	5/16	5.90	39.2	74.0	10.7	1.39	3.6	7.6*	68.4	1027	288	11.6	47.0	186	23.4	25.2	1862	1200	17.0	59
T-9	5/17	5.81	35.8	64.6	15.9	1.48	5.7	12.8*	56.3	1071	338	6.8	90.7	97.5	29.8	24.0	1453	1200	25.9	78
T-10	5/19	5.99	36.9	69.4	14.5	0.96	3.4	9.3*	62.9	1027	315	10.4	71.2	0	28.7	30.7	2248	1200	36.7	100

* Includes scrubber return and stack loss (Ref. Sec. 6)

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6. ANALYSIS OF TEST RESULTS

First Test Series

Normal Operation -- 100% Oil-Firing

The test indicated the variable and subjective nature of operator control. Specific energy consumption on an LHV basis varies between 11.9 to 14.3 GJ/tonne CaO (10.2 to 12.4 x 10^6 Btu/ton). Lime availabilities were lower than expected and corresponding carbonate remainders were low, of the order 0 - 1.8%. This indicated overburning of the lime. Of particular concern were the high oxygen levels, indicating inefficient combustion.

Baseline Operation -- 100% Oil-Firing

In this test of 37 hours duration, lower oxygen levels were maintained in the kiln by a combination of lower draft and primary air flow.

The oil flow was reduced without affecting product temperature and the exit gas temperatures were reduced on average. High availabilities indicated near optimal operation. Specific energy consumption on an LHV basis averaged 11.7 GJ/tonne lime product (10.1 x 10^6 Btu/ton).

Higher mud feed rates, less air leakage and further experimentation would have reduced oxygen levels to the 2% level. However, feed was limited by the capacity of the feed screw and air ingress could not easily be reduced. Accordingly, the above conditions were accepted as practical "baseline" for comparison purposes.

The baseline operation, as defined in Table 7 and Fig. 17, was modelled on the Lime Kiln Simulation resulting in the printout of parameter profiles given in Fig. 20 together with a typical baseline shell scan.

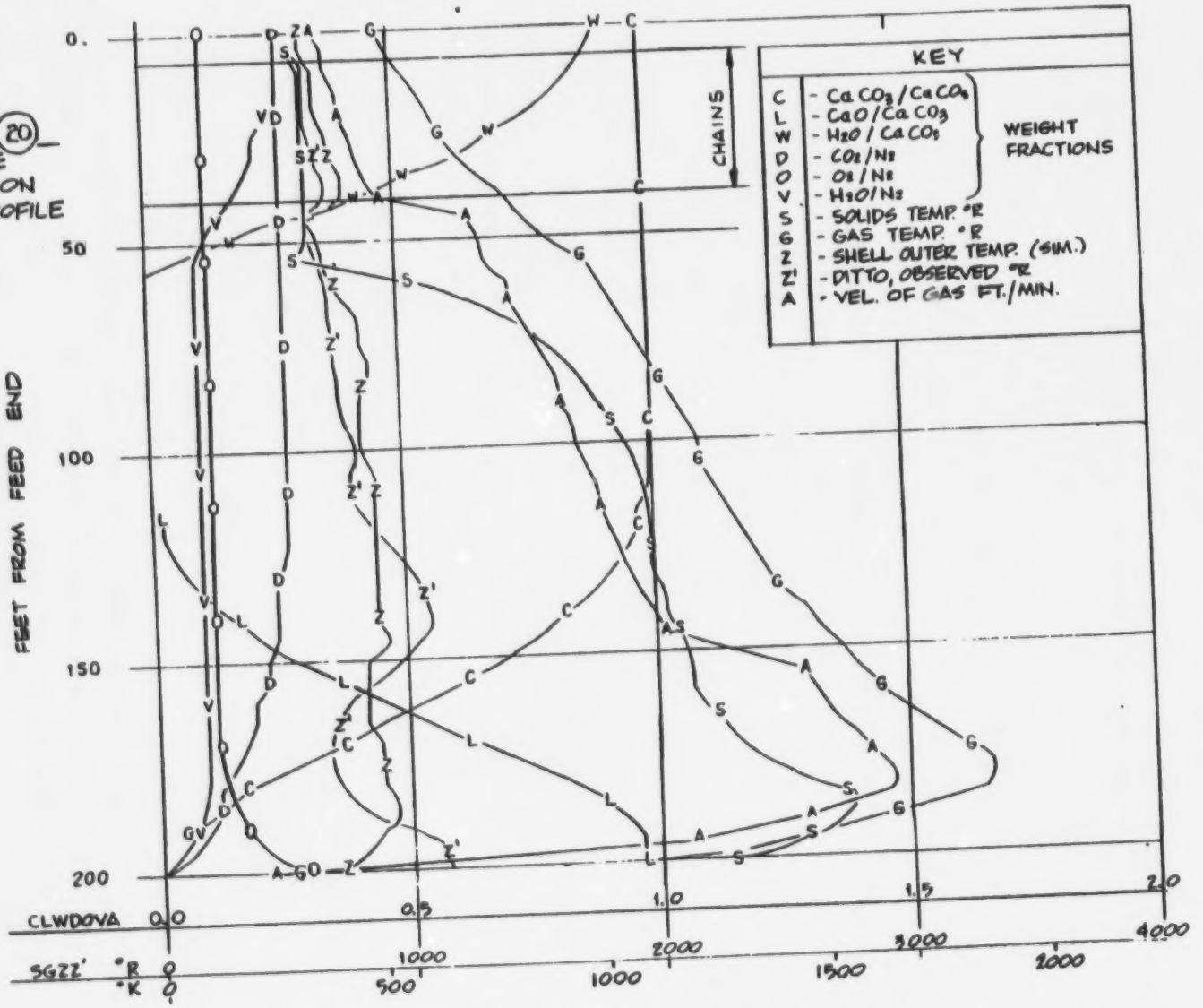
50% Oil Substitution

Upon introduction of Wet-Cell gases, the gas temperature profile changed rapidly, with a corresponding effect on product temperature, which reduced from 1200° C. (2200° F.) to a maximum of 1150° C. (2102° F.) and an average of around 1060° C. (1940° F.).

Gas exit temperature rose and it was necessary to maintain water on the feed screw in order not to exceed material limits. Temperatures were maintained in the region of 340° C. (644° F.). Oxygen levels of 7.5%, when compared to Wet-Cell oxygen levels of around 3%, indicate that ingress air volume was substantial. Since this represents an appreciable thermal load and exit temperatures were high, it is not surprising that the specific energy approximately doubled from 11.7 GJ/tonne CaO (10.1×10^6 Btu/ton CaO) to 25.4 GJ/tonne CaO (21.9×10^6 Btu/ton CaO). Specific energy allocated to source indicated that 6.86 GJ/tonne CaO (5.91×10^6 Btu/ton CaO) came from oil firing and 18.54 GJ/tonne CaO (16.0×10^6 Btu/ton CaO) came from the Wet-Cell. Comparison to baseline indicated that 1.0 oil thermal unit was replaced by 2.6 Wet-Cell thermal units. The replacement ratio and thermal efficiency are discussed further in conjunction with the Second Test Series.

Availabilities remained high and compared favourably to baseline conditions. Lime was qualitatively designated

FIGURE (20)
BASELINE SIMULATION DATA PROFILE



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"soft burned" and carbonate levels were higher than baseline. No settling problems were reported.

The trial duration was 78 hours, including an initial 20 hours at 25% oil substitution. Throughout that period, stable operation was maintained with minimal operator supervision, together with satisfactory calcination. A short interruption of output due to a fuel feed problem was handled without loss of production by increasing oil flow. The measures adopted in the kiln hood appeared to be effective.

Second Test Series

The second test series demonstrated calcination at all levels of oil substitution up to 100%. Satisfactory calcination (availabilities greater than 90%) was demonstrated at 78% oil substitution. The product lime had a characteristic brownish colour, but no settling problems were reported. Results are summarized in Table 8.

Replacement Ratio

In order to compare the performance of the system with dual-firing, to the 100% oil-fired baseline condition, it is necessary to consider energy distribution and efficiency of the system. An indication of the latter compared to baseline conditions is the replacement ratio. This is defined as:

$$\text{Replacement ratio, } R = \frac{Q_{wc}}{Q_{ob} - Q_{ot}} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (26)$$

where Q_{wc} = Wet-Cell heat supplied

Q_o = Oil heat supplied

TABLE 8 SECOND TEST SERIES -- DATA SUMMARY

Test	Date	Feed to Filter			Dust Loss (Scrubber)				Net Feed	Temperature		% O_2 By Vol.	Avail. %	Oil Fuel kg/h	CaO tPD	Specific Energy (LHV) GJ/t	Screw Water Added kg/h	Wet-Cell		% Oil Sub
										Product	Kiln Gas Out							Gas Out °C	Output GJ/h	
		m^3/h	% Solids	tPD	m^3/h	% Solids	tPD	% Loss		°C	°C									
Baseline	--	5.90	40.1	76.3	--	--	6.5	8.6	69.8			6.0	92.0	453	37.3	11.7	0			0
T-1	5/05	5.99	38.9	74.4	18.2	0.50	2.2	4.0*	71.4	1399	215	2.0	94.2	374	38.6	9.42	0			0
T-2	5/06	5.99	31.4	56.4	19.1	1.17	5.4	11.0*	50.2	1415	315	10.2	94.7	374	27.3	13.3	0			0
T-3	5/07	5.81	40.1	75.1	16.6	4.40	18.0	26.6*	55.1	982	354	9.4	49.2	235	19.5	32.4	435	1120	16.8	48
T-4	5/09	5.99	40.3	77.9	10.2	2.21	5.5	9.7*	70.3	1171	332	6.0	93.0	235	37.8	21.9	2021	1200	25.0	48
T-5	5/09	5.99	40.3	77.9	10.2	2.21	5.5	9.7*	70.3	1149	360	6.5	94.1	235	38.1	18.9	2021	1200	20.5	48
T-6	5/10	5.81	40.6	76.2	10.0	3.50	8.6	14.6*	65.1	1132	343	8.4	93.4	125	35.1	26.4	2748	1240	33.5	72
T-7	5/15	5.90	32.5	58.1	15.9	0.38	1.4	5.9*	54.7	1088	321	6.2	92.6	186	29.3	24.5	1476	1160	22.4	59
T-8	5/16	5.90	39.2	74.0	10.7	1.39	3.6	7.6*	68.4	1027	288	11.6	47.0	186	23.4	25.2	1862	1200	17.0	59
T-9	5/17	5.81	35.8	64.6	15.9	1.48	5.7	12.8*	56.3	1071	338	6.8	90.7	97.5	29.8	24.0	1453	1200	25.9	78
T-10	5/19	5.99	36.9	69.4	14.5	0.96	3.4	9.3*	62.9	1027	315	10.4	71.2	0	28.7	30.7	2248	1200	36.7	100

* Includes scrubber return and stack loss (Ref. Sec. 6)

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and subscripts, t = test
b = baseline
wc = Wet-Cell
o = oil

Thus R is the ratio of wood thermal units supplied per oil thermal unit replaced. For a given production level, a value of R = 1 represents a conversion in which the thermal efficiency is equal to that of the baseline. Similarly, higher values of R represent higher specific energy requirements and lower efficiency than baseline, assuming constant production.

With operator control, in the first test series, the kiln showed specific energy values higher than baseline, i.e., 12 - 15 GJ/tonne CaO. In order to assist in validation of the computer simulation, Test 1 and 2 of the second series were executed with low and high oxygen respectively. These showed specific energy values of 8.5 and 13.3 GJ/tonne CaO, bracketing the baseline value of 11.7 GJ/tonne CaO, which was therefore considered representative of the normal well-tuned operation.

Heat input, replacement ratio and specific energy values for the second test series are shown in Table 9.

The tests may be interpreted in groups of similar substitution, i.e., Tests 3, 4 and 5 at 48%, Tests 7 and 8 at 59%, and Tests 6 and 9 at 72% and 78% substitution respectively.

Test 3 showed poor production due to low Wet-Cell input combined with high dust loss (Table 8). The replacement ratio

TEST	LIME		OIL HEAT IN	WET-CELL HEAT IN	TOTAL HEAT IN	SPECIFIC ENERGY	REPLACEMENT RATIO	SUBSTITUTION
	tPD	AVAIL	GJ/h			GJ/t		%
Base	37.3	92.0	18.4	0	18.4	11.8		0
1	38.6	94.2	15.2	0	15.2	9.4		0
2	27.3	94.7	15.2	0	15.2	13.4		0
3	19.5	49.2	9.52	16.8	26.3	32.4	1.8	48
4	37.8	93.0	9.52	25.0	34.5	21.9	2.8	48
5	38.1	94.1	9.52	20.5	30.0	18.9	2.3	48
6	35.1	93.4	5.05	33.6	38.6	26.4	2.5	72
7	29.3	92.6	7.54	22.4	29.9	24.5	2.0	59
8	23.4	47.0	7.54	17.0	24.5	25.1	1.6	59
9	29.8	90.7	3.95	26.0	29.9	24.1	1.8	78
10	28.7	71.2	0	36.7	36.7	30.7	2.0	100

TABLE (9)

SECOND TEST SERIES -- HEAT INPUT AND REPLACEMENT RATIOS

of 1.89 (i.e., 1.89 wood Btu/oil Btu replaced), combined with low gas temperature (1120° C.), led to inadequate calcination.

Test 4 showed good production and high availability with increased Wet-Cell input (R=2.81). The Wet-Cell input was reduced to a more optimal level in Test 5 (R=2.31) without affecting production, although air leakage was still high.

Test 7 (R=2.05) showed good calcination but a small drop in production due to low % solids in the feed. In Test 8, (R=1.56) the Wet-Cell input was too low. Combined with continued high air leakage, this resulted in poor calcination and low production.

Comparison of Test 6 (R=2.51) and Test 9 (R=1.79) appeared to show higher substitution in the latter with reduced Wet-Cell input. However, reference to Table 8 highlights the reduced effective feed rate of Test 9, which therefore required less energy.

Test 10 (R=1.99) at 100% oil substitution showed unsatisfactory calcination under the prevailing conditions, and a subsequent test run produced availabilities less than 71.2. However, Test 9 availabilities were acceptable at reduced feed-rate.

This suggests that adequate availability could be maintained with near 100% wood-firing, if feed rate is reduced.

Energy Distribution and Losses

Estimates of energy distribution for Tests 5, 8, 9 and base-line and shown in Figures 21, 22, 23 and 24 respectively.

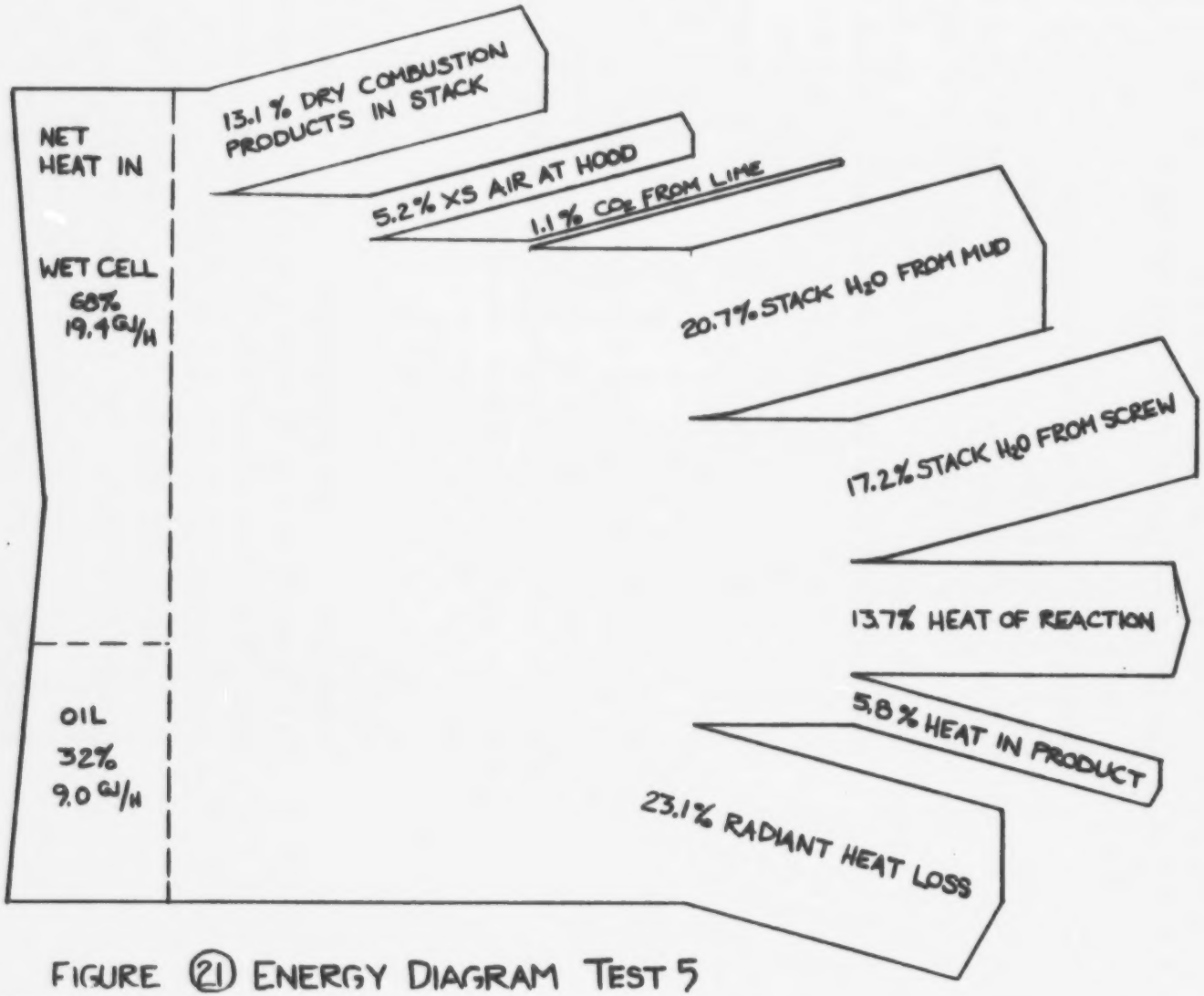

FIGURE (21) ENERGY DIAGRAM TEST 5

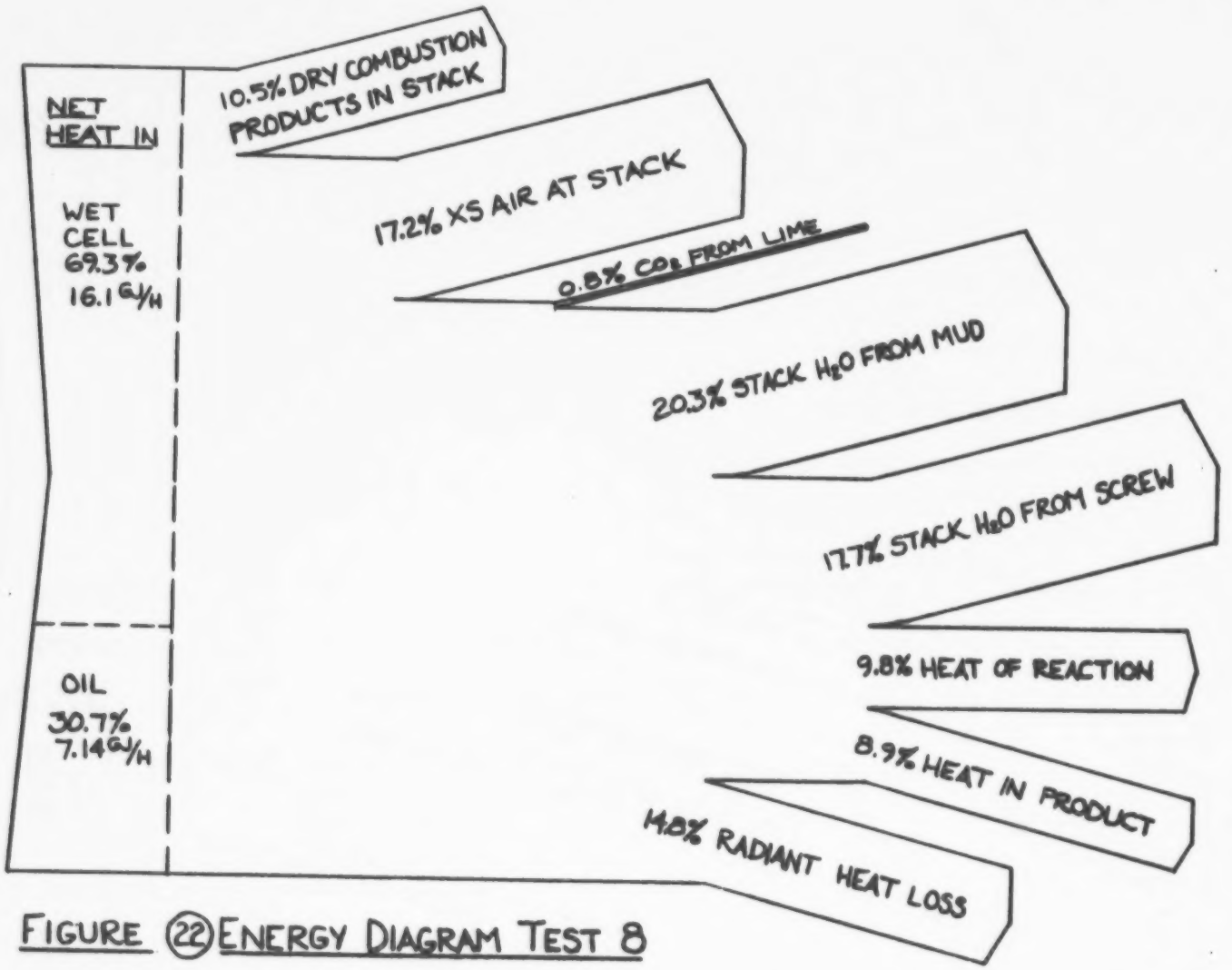
NET HEAT IN
WET CELL 69.3% 16.1 GJ/H
OIL 30.7% 7.14 GJ/H
10.5% DRY COMBUSTION PRODUCTS IN STACK
17.2% XS AIR AT STACK
0.8% CO2 FROM LIME
20.3% STACK H2O FROM MUD
17.7% STACK H2O FROM SCREW
9.8% HEAT OF REACTION
8.9% HEAT IN PRODUCT
14.8% RADIANT HEAT LOSS

FIGURE (22) ENERGY DIAGRAM TEST 8

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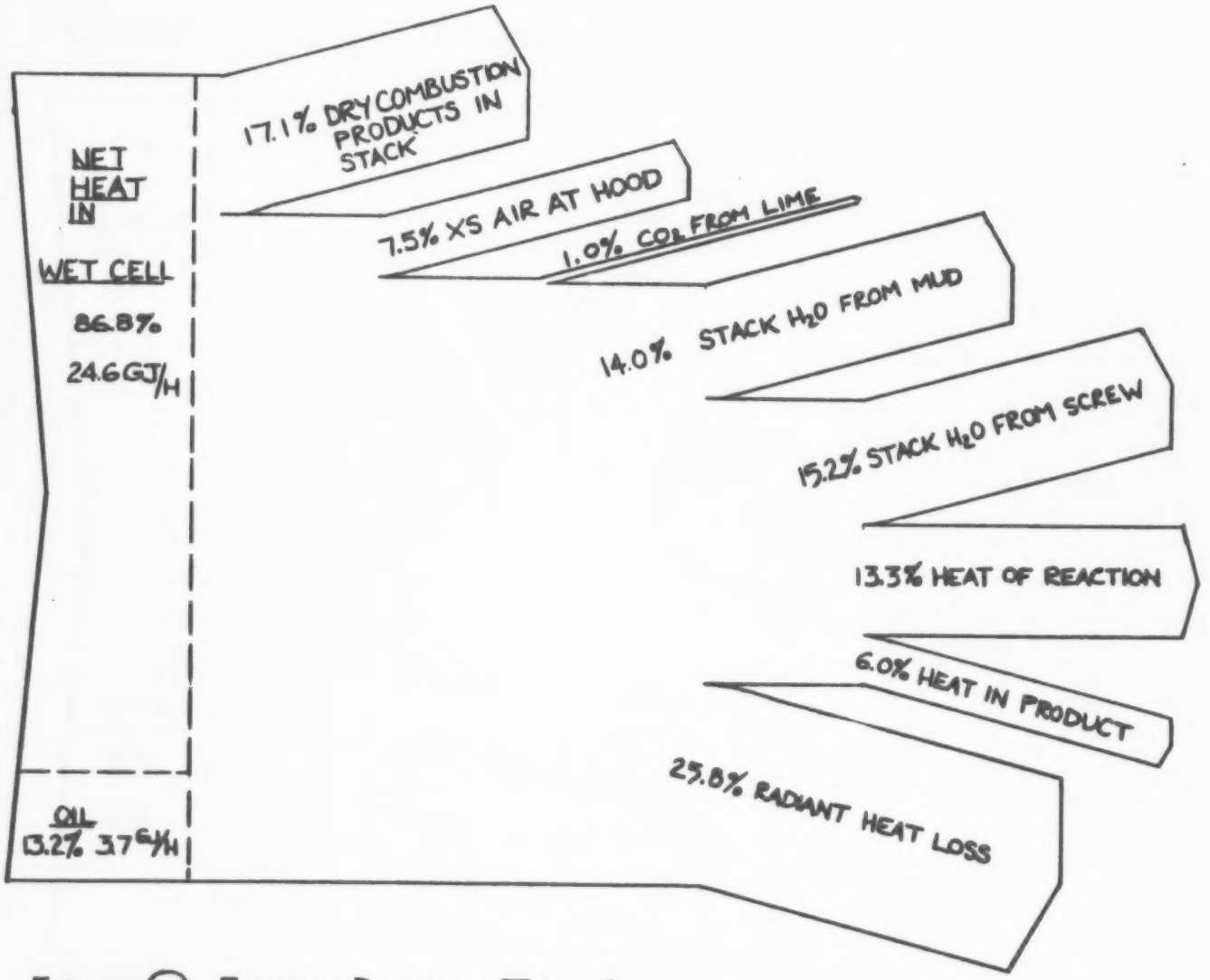

FIGURE (23) ENERGY DIAGRAM TEST 9.

LAMB-CARDATE INDUSTRIES LIMITED

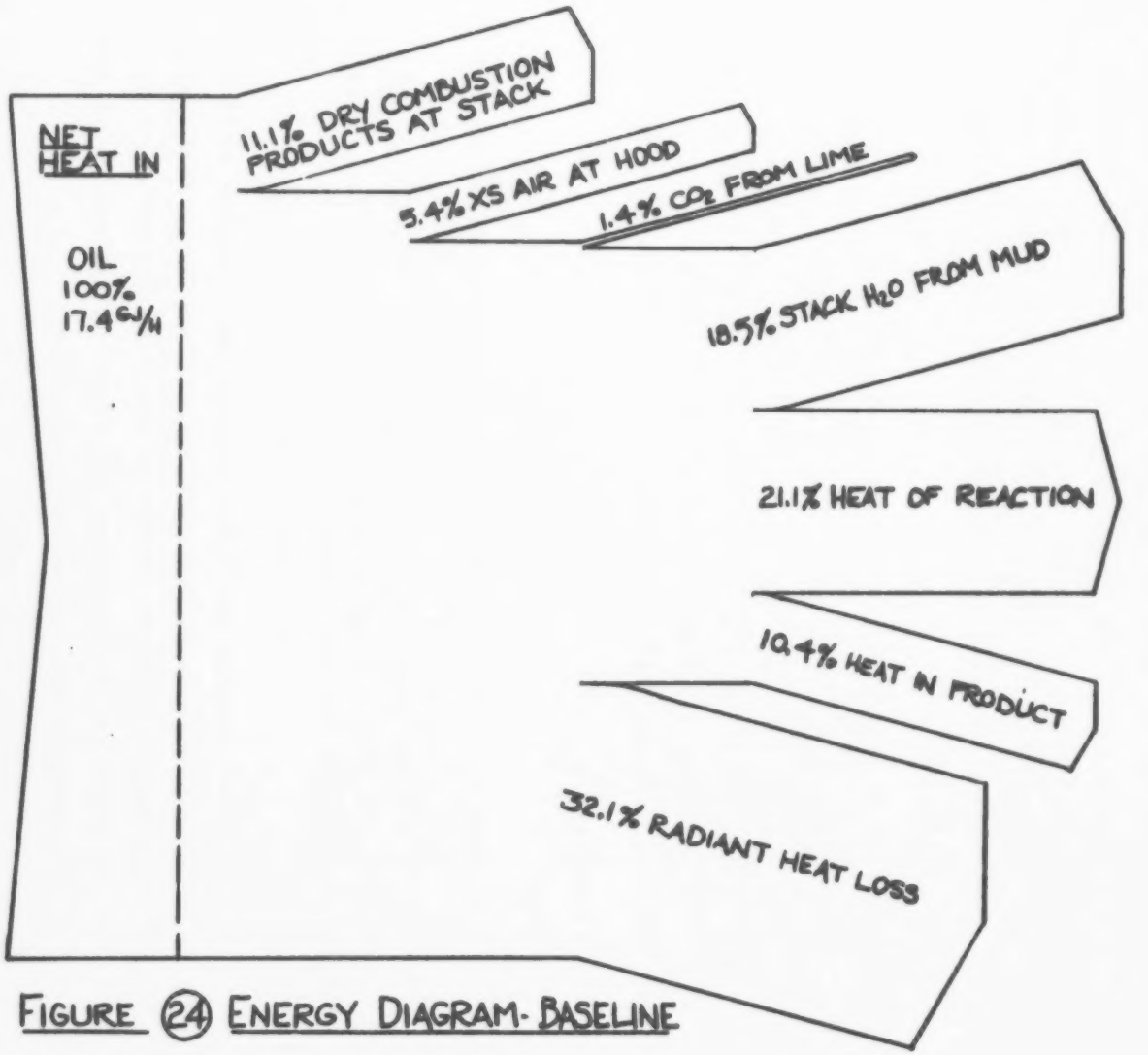

FIGURE (24) ENERGY DIAGRAM - BASELINE

The stack heat loss associated with the water vapour in the combustion gases is not shown. This is already accounted for by using the net or lower heating values for the oil and wood fuel.

Nevertheless, heat is required to evaporate water from the mud, and raise the water vapour to the kiln exit temperature. This heat use may be divided into (a) normal mud drying, and, (b) evaporation of water added to the feed screw to suppress exit gas temperatures as shown in the diagrams.

It is important to recognize that the suppression of exit gas temperature by this means does not impose an extra thermal load on a dual-fuel-fired system; examination of the simulation profiles in Section 7 shows that mud drying is accomplished well within the chain section and calcination is not inhibited. However, it is a loss which must be added to the dry combustion product and excess air loss for comparison to the baseline case.

Thus, for Tests 5, 8 and 9, the major stack losses (combustion products + excess air + screw water) are 35.5, 45.4 and 39.8% of net heat input respectively. The equivalent baseline loss (combustion products + excess air) is 16.5% only.

The stack losses are higher because heat transfer is diminished by low temperature differences between gas and product or wall. Increased convective heat transfer may occur as a result of increased mass velocity, but does not fully compensate for the reduced gas temperature.

The energy losses attributable to ingress air at the hood are shown. The mass flows in the system, calculated by the methods of Section 4 (Page 34) are shown in Table 10. The excess air cools the Wet-Cell gases and reduces the gas-to-solids temperature difference. The excess air occurs as a result of the following:

(a) The secondary air nose ring seal system which operated throughout the tests.

(b) The high primary air on the oil gun. While this reduced efficiency, it improved flame stability and cooled the oil gun, preventing the occurence of thermal distortion. It was more difficult to maintain the small oil flame associated with high substitution so a trial and error procedure was followed to minimize primary air on each test.

(c) Air leaked into the kiln hood from the product chain trough.

Since the minimum air requirements for the system are determined by the oil gun, a primary objective of conversion design would be to limit all other air leakage. Items (a), (b) and (c) suggest design refinements which are discussed in Section 9.

TABLE (10) SECOND TEST SERIES -- GAS MASS FLOW

	Wet-Cell	Oil	Air Into Hood	CO_2 Off Mud	Water On Screw	Water In Mud	Water In Air	Wet-Cell O_2	Kiln O_2 Out	Total Flow
	kg/h	kg/h	kg/h	kg/h	kg/h	kg/h	kg/h	% Wt	% Wt	kg/h
Base		453	10624	1218	0	1683	126	--	7.1	14104
T-1		374	6330	1274	0	1898	82	--	2.4	9928
T-2		374	13520	906	0	1239	176	--	12.0	16215
T-3	10830	235	15350	662	435	1720	199	4.2	11.1	29430
T-4	15960	235	11360	1271	2021	2436	148	3.1	7.1	33430
T-5	12820	235	12080	1280	2021	2436	157	2.6	7.7	31000
T-6	21170	125	18660	1191	2748	2136	242	3.1	9.9	46270
T-7	15080	186	7320	790	1476	990	95	5.3	7.3	25940
T-8	10680	186	25850	992	1862	2135	336	3.2	13.7	42000
T-9	16600	97.5	9066	1022	1453	1338	216	3.6	8.0	29800
T-10	22630	0	24530	980	2248	1561	318	3.0	12.3	52270

Stack and Scrubber Loss

A total of eight emissions tests were carried out throughout the testing period. Summarized data are shown in Table 11. No obvious correlation is apparent between mass flow and dust loss.

This is somewhat surprising in view of the increased gas/bed mixing suggested on Page 80. However, in order to reach the stack, particulate must escape collection by:

(a) the chain section, where collection efficiency is likely to improve with increased mass flow
(b) the scrubber
(c) centrifugal action in the exit gas ducting.

Dust sludge was observed to build up in the sumps below the shower chambers and regular clean-up of the scrubber nozzles was required. The limited number of the emission tests, and the variable performance if the equipment at the kiln exit, has not permitted identification of a clean trend.

Despite this, it was still necessary to allow for stack loss in the calculation of net feed input. The following values which are approximately consistent with those found in the emissions tests, were used:

50% substitution	83 kg/h
75% substitution	106 kg/h
100% substitution	128 kg/h

TABLE 11 GAS MASS FLOW VS. STACK DUST LOSS

Gas Mass Flow kg/h	% Substitution	Dry Solids Feed kg/h	Scrubber Loss kg/h	Scrubber Loss % Feed	Stack Loss kg/h	Stack Loss % Feed	Total Loss kg/h	Total Loss % Feed
15120	0	3126	480	15.3	44	1.4	524	16.7
15560	0	3315	211	6.4	36.6	1.1	248	7.5
25814	48	3194	298	9.3	179	5.6	478	14.9
28260	24	2998	434	14.5	108	3.6	542	18.1
27700	78	2619	229	8.8	82	3.1	311	11.9
31220	72	3421	155	4.5	40	1.2	195	5.7
39490	100	2521	232	9.2	156	6.2	389	15.4
41291	100	3364	268	7.9	93	2.8	361	10.7

The scrubber recycle loss generally stayed at 5 to 10% of filtered feed solids. The stack loss with oil-only conditions is limited to 1 to 1.5% of filtered feed solids. Above 73% substitution, the average stack loss was 4% of filtered feed solids while scrubber recycle losses remain around the 7% level.

Test Conclusions

A primary objective of the project was the evaluation of the performance limits for dual-fuel operation. In this connection, the results of the test series may be summarized as follows:

Calcination: The tests demonstrated calcination with up to 100% substitution with wood heat. Adequate availability, greater than 90% was maintained at substitution levels of 48 to 78% on this system.

Production: the dual-firing tests were carried out with lime production up to and slightly beyond baseline level. Data acquired reflected the impact on production of gas temperatures, heat input, percent solids in the mud feed and dust loss.

Energy Use: Energy input/tonne product was higher with wood heat than with 100% oil-firing at baseline. The necessity to compensate for reduced gas temperatures on wood heat, by increasing gas flow, leads to higher losses. Cooling air at the hood was identified as one of the causes of reduced efficiency. Air ingress can be minimized in conversion design. The major cause of reduced thermal efficiency is reduced heat transfer. This results in high exit gas temperatures which must be restrained by adding water to the feed screw.

The test series provided a data base for evaluation of the computer simulation.

7. COMPUTER SIMULATION

As detailed in Section 4, validation of the computer simulation is established when simulated profiles are compared successfully with actual kiln data for steady-state operating conditions, and additionally, when the model parameters can be varied to match kiln test data with the changed conditions.

Difficulties were experienced in making the model fit observed data for dual-firing, and in particular, the model showed reduced performance as a result of reduced heat transfer. This necessitated some attention to the assumptions of the model design in particular, consideration of the convective heat transfer coefficients and of radial temperature profiles.

Heat Transfer and Simulation Validation

A basic analysis of heat flow paths in a rotary kiln is provided by Brimacombe and Watkinson (8), Fig. 25, whereby:

$$\left(\frac{q}{L}\right)_{g \longrightarrow w+s} = \left(\frac{q}{L}\right)_{gw} + \left(\frac{q}{L}\right)_{gs} \quad \text{.....(28)}$$

and

$$\left(\frac{q}{L}\right)_{w+g \longrightarrow s} = \left(\frac{q}{L}\right)_{ws} + \left(\frac{q}{L}\right)_{gs} \quad \text{.....(29)}$$

and a wall heat balance yields:

$$\left(\frac{q}{L}\right)_{a} = \left(\frac{q}{L}\right)_{gw} - \left(\frac{q}{L}\right)_{ws} \quad \text{.....(30)}$$

Where $\frac{q}{L}$ = heat/unit time, and L = incremental length, subscripts:

- w = wall
- s = solids
- g = gas
- a = atmosphere
- i = inside shell
- o = outside shell

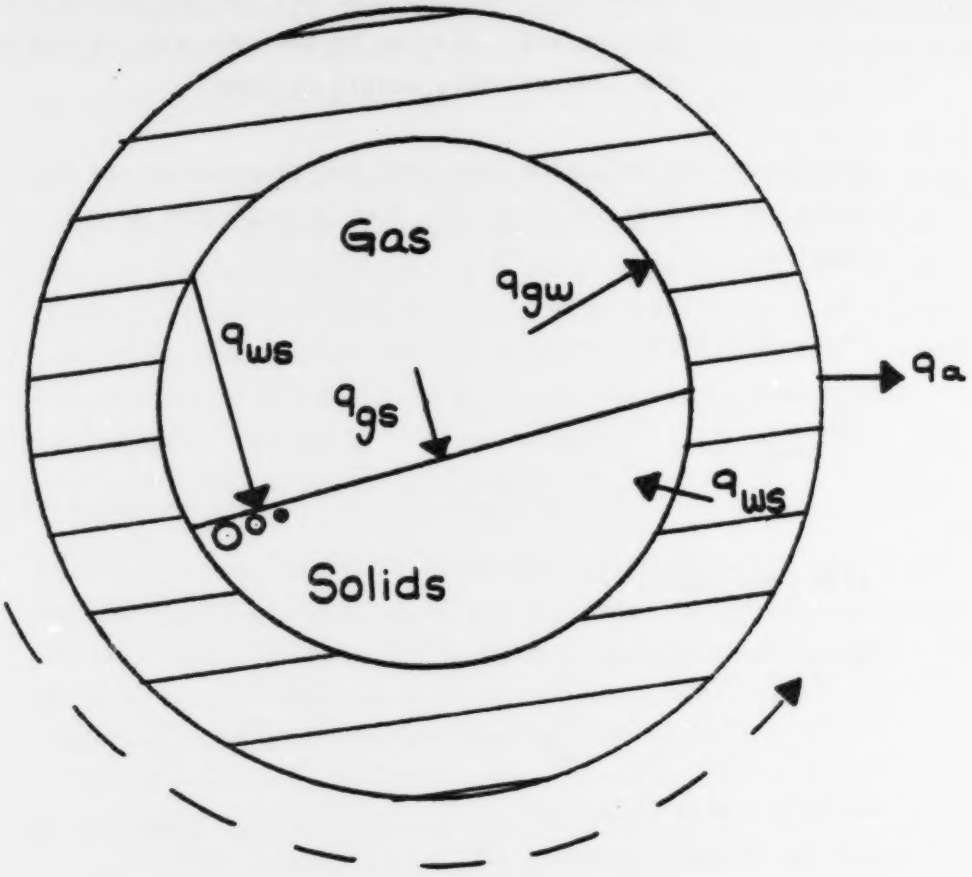

FIG (25) HEAT FLOW PATHS IN A ROTARY KILN.

Heat transfer modes ascribed to these terms are:

q_{gw} -- radiation plus convection, heat/unit time

q_{gs}

q_{ws} -- radiation from uncovered wall to top of bed plus conduction and radiation to the underside of the bed from the covered wall, heat/unit time.

Brimacombe and Watkinson note that bed temperature and wall temperature are similar and that $(q_L)_{ws}$ can be neglected as a first approximation.

Thus, (29) reduces to:

$$(q_L)_{(w+g) \dashrightarrow s} = (q_L)_{gs}$$

$$= G_s \frac{dT_s}{dx} [C_{ps}] \quad \ldots\ldots(31)$$

Also from (28):

$$(q_L)_{g \dashrightarrow (w+s)} = (q_L)_{ws} + (q_L)_s$$

$$= G_g \frac{dT_g}{dx} [C_{pg}] \quad \ldots\ldots(32)$$

where G = mass velocity,

and C_p = specific heat,

$(q_L)_a$ is fixed if the simulation outside wall temperature scan matches the observed scan. G_s and G_g are measurable input quantities and the values of C_p can be allocated with accuracy. Thus $\frac{dT_s}{dx}$ and $\frac{Dt_g}{dx}$ can be derived from (31) and (32) and the model is validated for steady-state conditions.

Convective Heat Transfer

The simulation model includes convective and radiative components of heat transfer coefficients which would yield q_{gw}, q_{gs}, and q_{ws}.

However, the convective component remains fixed while the radiative component changes with temperature as the iteration proceeds. For the baseline case, Boeing experience has confirmed convective heat transfer coefficients of the order 73 to 360 kJ/m²h°C., (3.6 to 18 Btu/ft²h°F.) depending on the zone within the kiln, for both gas to wall and gas to solids.

For conventional lime kiln analysis, this approach would be valid since convective heat transfer depends upon kiln geometry, mass velocity and gas properties which would normally not vary appreciably. In the case of dual-firing with a Wet-Cell, mass velocities vary substantially and the effect on convective heat transfer must be evaluated. An increase in the convective heat transfer may partially compensate for the reduction in radiative heat transfer associated with Wet-Cell operation.

It was therefore decided to vary the model convection heat transfer coefficients according to the Reynolds Analogy of heat transfer to momentum exchange. The Analogy enables power law relationships to be developed between dimensionless model reatios or numbers as follows:

$$Nu = C\ (Re)^{n}\ (Pr)^{n_2} \qquad (33)$$

Where:

Nu = Nusselt Number = $\frac{hD}{k}$

C = Constant

h = Heat transfer coefficient (Heat/area - time - °C)

Re = Gas Reynolds Number, dimensionless = $\frac{G_g D_i}{\mu_g}$

Pr = Gas Prandtl Number, dimensionless = $\frac{C_{pg}\mu_g}{k_g}$

D = Representative diameter

k_g = Gas conductivity (Heat/area - time - degree-unit length)

μ_g = Gas viscosity (Force/unit length - time)

Since flow was fully turbulent, with Re in the range 20,000 to 200,000, the empirical relationship:

$$h_{gw} = 0.05\ G_g^{0.67} \quad \text{..... (35)}$$

may be used to pro-rate h_{gw} from baseline conditions (9).

However, for h_{gs}, Brimacombe and Watkinson (8) comment on the difficulty of selecting a useful correlation for heat transfer from a moving gas stream to a moving solids bed. The equation for boundary layer flow over a flat plate:

$$h_{gw} \propto \frac{0.332\ k_g}{L-x} (Re_x^{0.5}\ Pr_g^{0.33})$$

where L = length, and x = distance gave low values. Though their experiments took place in a pilot kiln at low Reynolds number, it was apparent that the gas to solid heat transfer was an order of magnitude larger than the gas to wall transfer.

Normal practice for 100% oil-firing is to use values higher than those given by Equation (35). This is not surprising since in most industrial cases, the bed remains in a rolling mode, suggesting a high degree of permeation of the burden by the heating gas and continuous disturbance of the boundary layer.

The Reynolds/Nusselt equation for gas fluidized beds is a representation of extreme permeation and gives high values for the heat transfer coefficient. Frantz ex Perry (10) proposed:

$$\frac{hD_p}{k_g} = 0.015\ \left(\frac{D_p G_g}{\nu_g}\right)^{1.6} (Pr)^{0.67} \quad \text{..... (36)}$$

where subscript $_p$ refers to the particle in the bed.

For turbulent flow inside long tubes, $n_2 = 0.4$ is in general use (11). For n_1, the value for this application must lie between $n_1 = 0.5$ from Equation (35) and $n_1 = 1.6$ from Equation (36). Though fluidization of the bed does not occur, it is reasonable to assume a large exchange of momentum at the rolling bed interface. A value of $n_1 = 1.2$ was selected, and the equation:

$$\frac{h_{gs} D}{k_g} = 0.007 (Re)^{1.2}(Pr)^{0.4} \quad \text{.....(37)}$$

was used to relate convective coefficients to mass flow and gas transport properties.

The simulation fit (discussed hereafter) thus provided a useful insight into the heat transfer processes involved and it is suggested that convective transfer, in particular from gas to bed, contributes substantially to the heat balance and partially compensates for reduced radiation as oil (or natural gas) substitution increases. Equation (37) was used to provide optimization data of Section 8. Confirmation of C_1, n_1, and n_2 by rigorous means is recommended. It is important to realize that convective heat transfer can be enhanced by mechanical means to improve overall thermal performance.

The above reasoning applies to Zones 2, 3 and 4 only, since Zone 1 is a chain section and heat transfer is substantially greater, being a combination of forced convection and film evaporation. Empirical model coefficients for Zone 4 were an order of magnitude higher than Zones 2, 3 and 4.

Radial Temperature Profile

In order to represent radiant heat transfer at the firing end of the kiln, consideration must be given to the radial temperature profile. With 100% oil-firing, the oil flame occupies a large

volume of the kiln and the bulk gas temperature, developed from the combustion product enthalpy can be used to derive the radiant heat transfer with accuracy. With dual-firing, the oil flame is diminished in volume, but its core temperature remains high and it continues to radiate heat. However, without a radial temperature profile capability, the model does not develop the core temperature, but instead develops the bulk temperature of a mixture of oil and Wet-Cell combustion products, and leakage or excess air. The bulk temperature thus developed is substantially lower than the core temperature and the radiant capability of the reduced oil flame is not represented.

In order to compensate for this, it was necessary to inflate the value of bulk temperature to match the observed data. This inflation distorts the energy balance of the simulation, but does not invalidate the model as a predictive tool providing the degree of distortion is quantified.

Simulation Profiles

Simulated and observed data for the range of Tests are presented in Table 11. Parameter profiles are presented in Figures 26 to 33 for Tests 3 to 10 respectively. The profiles include observed shell scans for comparison with simulated outer wall temperatures.

Examination of the profiles on a Test-by-Test basis, and comparing them to the Baseline case (Fig. 20) provides the following insights:

Test 3 (48% substitution)

Calcination starts at 125-feet compared to 110-feet. Mud is dried much earlier and solids temperature rises within chain section. Gas temperature profile is flat and gas to

TEST NO.		MUD m^3/h	% SOLIDS	OIL kg/h	% O_2 (Vol.)	PRODUCT TEMP °C	EXIT GAS TEMP °C	PRODUCTION		
								AVAIL %	DUST %	NET tPD
BASE	T	6.0	40.1	453	6.0		293	92.0	8.6	37.3
	S	6.0	40.1	453	6.8	1100	260	92.6	8.6	37.5
1	T	6.0	38.9	374	2.0	1399	215	94.2	4.0	38.6
	S	6.0	38.9	374	4.0	1522	189	92.6	4.0	38.2
2	T	6.0	31.4	374	10.2	1415	315	94.7	11.0	27.3
	S	6.0	31.4	374	7.7	1208	264	88.6	11.0	26.2
3	T	5.8	40.1	235	9.4	982	354	49.2	26.6	19.5
	S	6.0	40.1	235	10.4	915	397	41.6	26.6	17.3
4	T	6.0	40.3	235	6.0	1171	332	93.0	9.7	37.8
	S	6.0	40.3	235	6.6	1091	358	93.4	9.7	37.9
5	T	6.0	40.3	235	6.5	1149	360	94.1	9.7	38.1
	S	6.0	40.3	235	7.2	1111	317	93.7	9.7	38.0
6	T	5.8	40.6	124	8.4	1132	343	93.4	14.6	35.1
	S	5.8	40.6	124	9.0	1125	468	92.0	14.6	34.8
7	T	5.9	32.5	186	6.2	1088	321	92.6	5.9	29.3
	S	5.8	32.5	186	6.6	1173	335	94.5	5.9	29.6
8	T	5.9	39.2	186	11.6	1026	288	47.0	7.6	23.4
	S	5.9	39.2	186	12.5	983	417	47.8	7.6	23.7
9	T	5.8	35.8	97.5	6.8	1071	338	90.7	12.6	29.8
	S	5.8	35.8	97.2	7.2	1169	353	88.7	12.8	29.5
10	T	6.0	36.9	0	10.4	1027	315	71.2	9.3	28.7
	S	6.0	36.9	0	10.9	1027	499	62.9	9.3	26.4

T = Test
S = Simulation

TABLE (12)

SIMULATION VALIDATION COMPARISON

solids temperature difference is around 100° C. Final calcination is low but climbing. There is no burning zone. Shell scan fit is excellent.

Test 4 (48% substitution)

Calcination starts at 75-feet. Mud is dried and solids temperature rises within chain section. Gas temperatures higher than Test 3 and gas to solids temperature difference is generally greater than 150° C. Calcination profile flattens out at approximately 180-feet, suggesting burning zone from 180 to 200-feet. Shell scan fit is good.

Test 5 (48% substitution)

Profiles are very similar to Test 4. Gas velocities are lower due to reduced Wet-Cell input. Shell scan fit is good.

Test 6 (72% substitution)

Calcination starts at 60-feet. Solids temperature rises within chain section. The gas temperature profile is very similar to Tests 4 and 5 though exit temperature is higher. Gas to solids temperature is around 150° C. Calcination profile rises continuously. Shell scan fit is very good.

Test 7 (59% substitution)

Calcination starts at 80-feet. Calcination profile flattens out at 180-feet. Shell scan fit is very good.

Test 8 (59% substitution)

Calcination does not start until 100-feet. The gas temperature profile is low and solids temperature profile

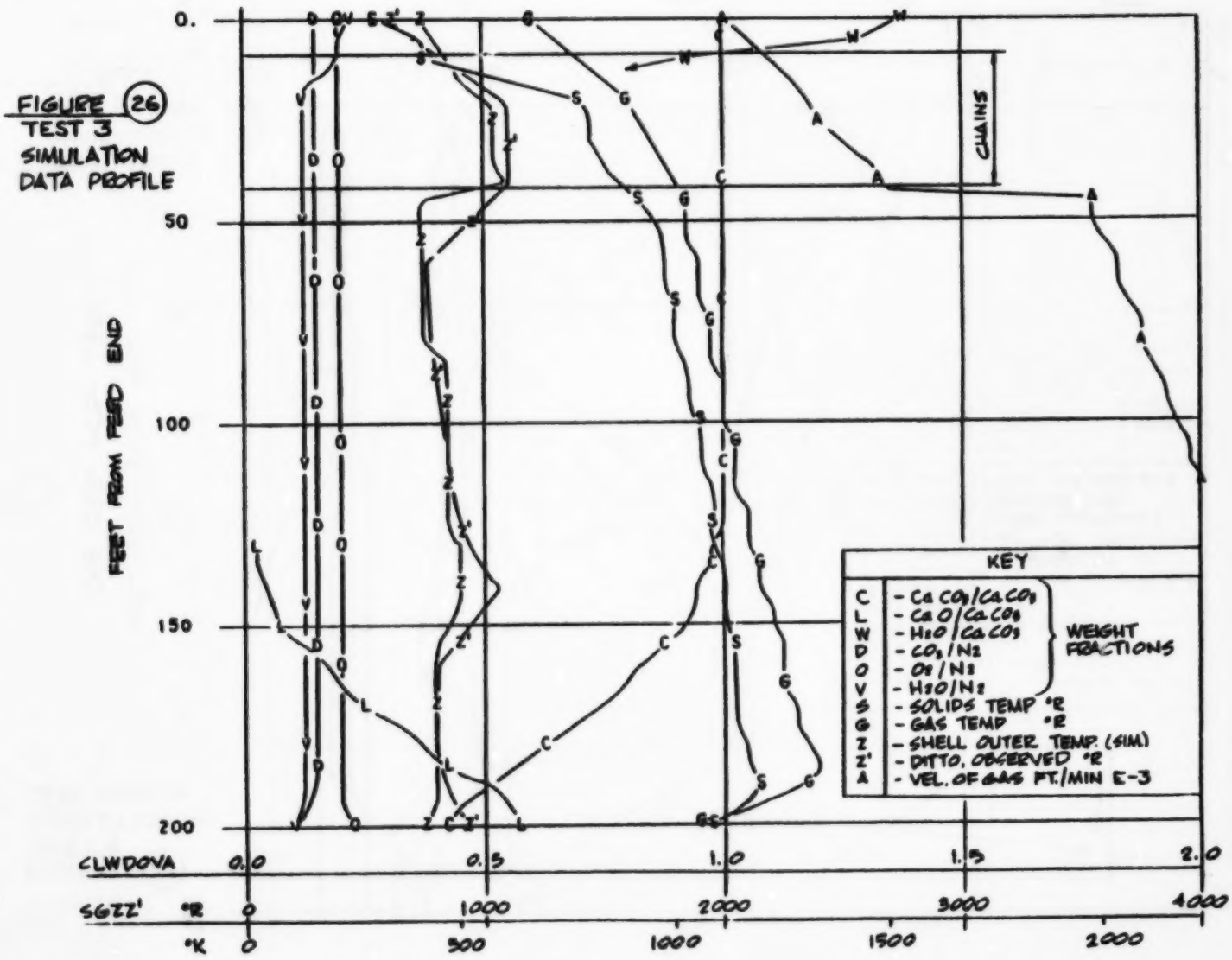

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FIGURE (27)
TEST 4
SIMULATION
DATA PROFILE

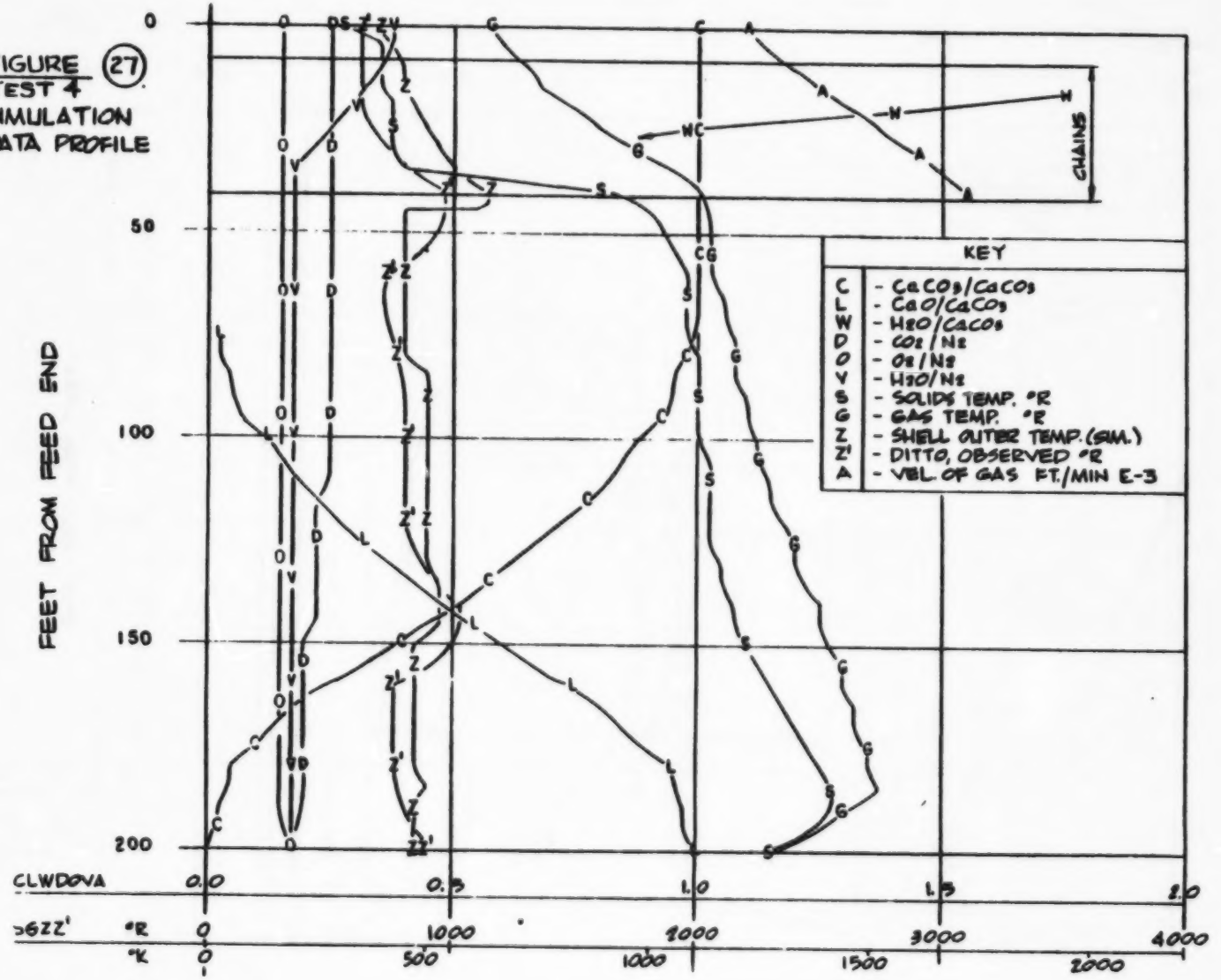

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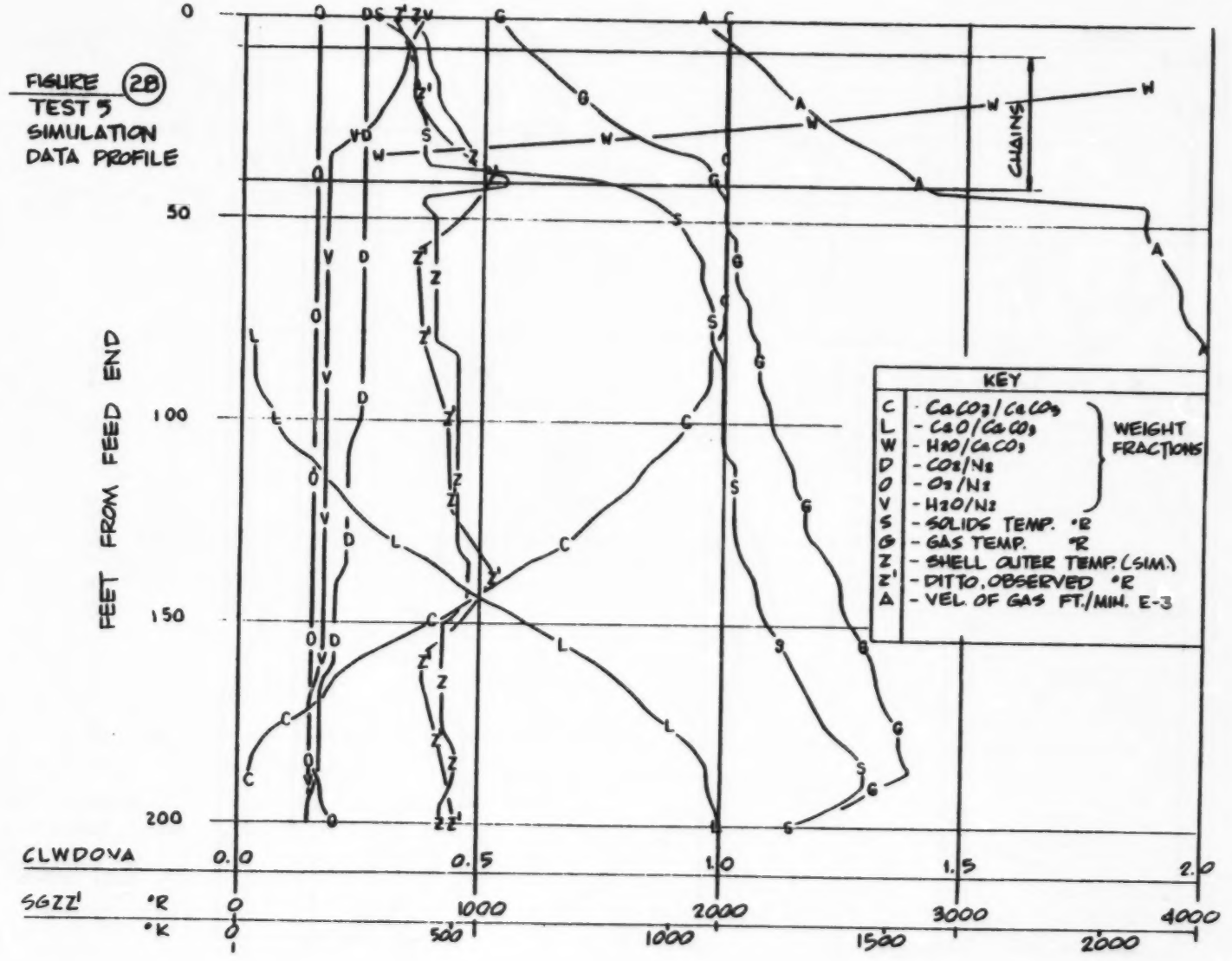
FIGURE 28
TEST 5
SIMULATION
DATA PROFILE



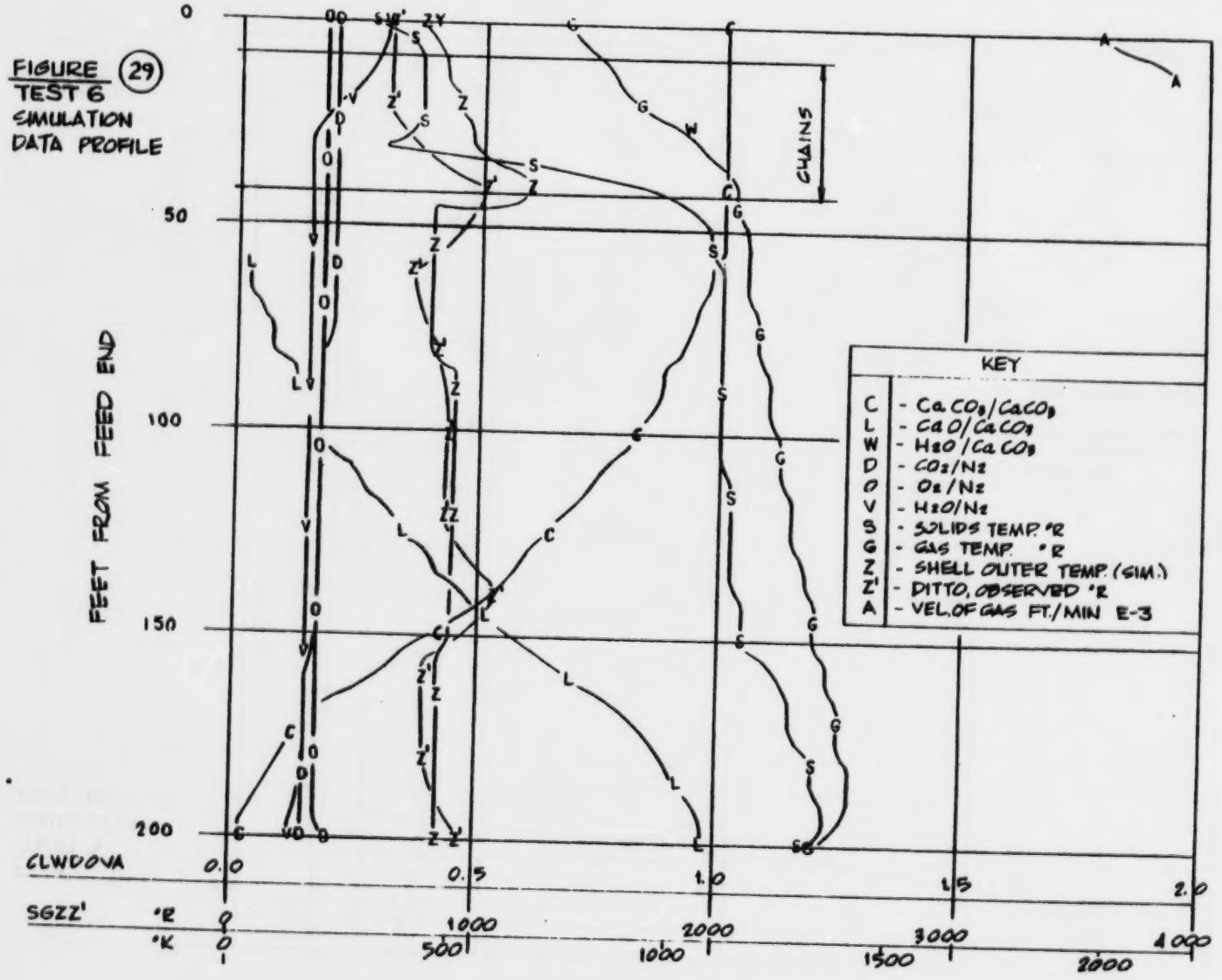


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FIGURE (30)
TEST 7
SIMULATION
DATA PROFILE

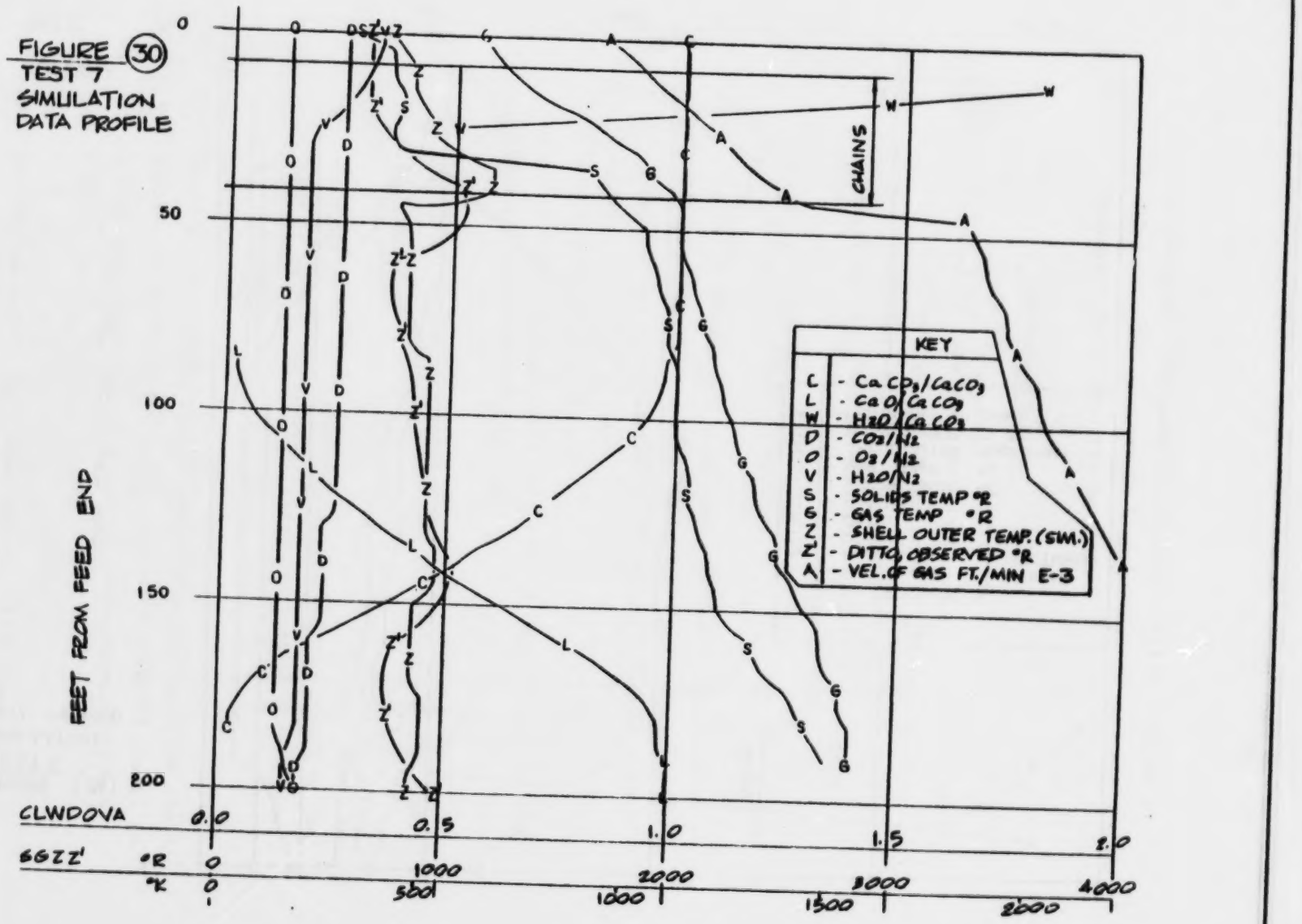
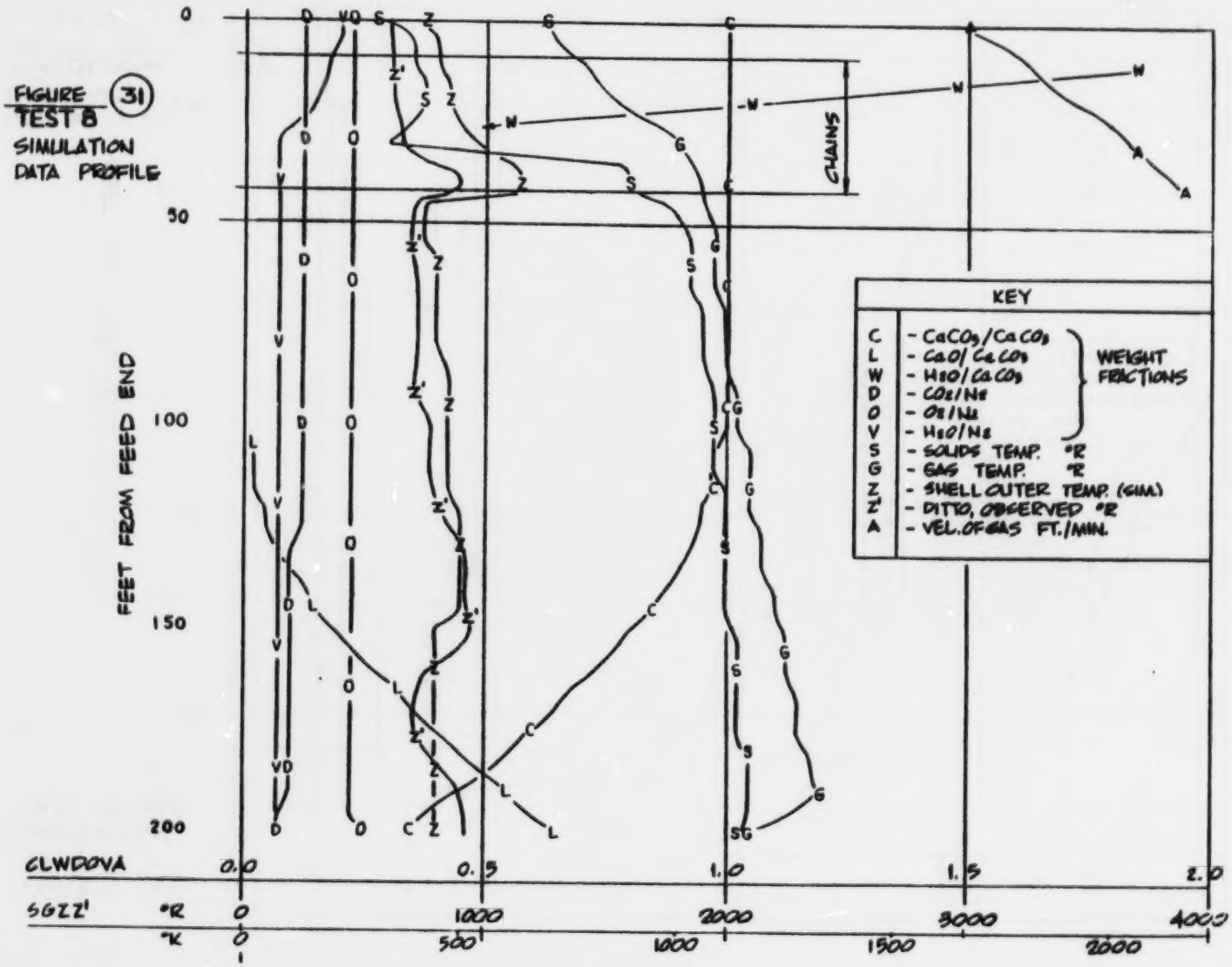


FIGURE 31
TEST 8
SIMULATION
DATA PROFILE



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corresponds. Calcination profile is still climbing at 200-feet. Gas velocity is high because of high air leakage. Shell scan fit is good.

Test 9 (78% substitution)

Calcination does not start until 110-feet but calcination profile climbs quickly to maximum at 190-feet despite low gas temperature profile. Good calcination is therefore attributed to low feed rate. Shell scan fit is good.

Test 10 (100% substitution)

Calcination does not start until 80-feet. Calcination profile is still rising at 200-feet. Gas velocities are high but gas temperatures are low. Shell scan fit is good.

Comparing these profiles with the Baseline case, (Fig. 20) the following observations may be made:

(a) The gas temperature profiles are relatively flat for dual-firing.

(b) Calcination is initiated earlier, at the mid-length of the kiln or before. With oil-firing, calcination occurs later, at 60 to 70% travel.

(c) Despite water on the feed screw, solids drying in the chain section is improved and solids temperature at 25% travel is higher than Baseline. Chain metal temperatures will be higher.

(d) Calcination occurs with lower gas to solids temperature difference, indicating that residence time is an important factor.

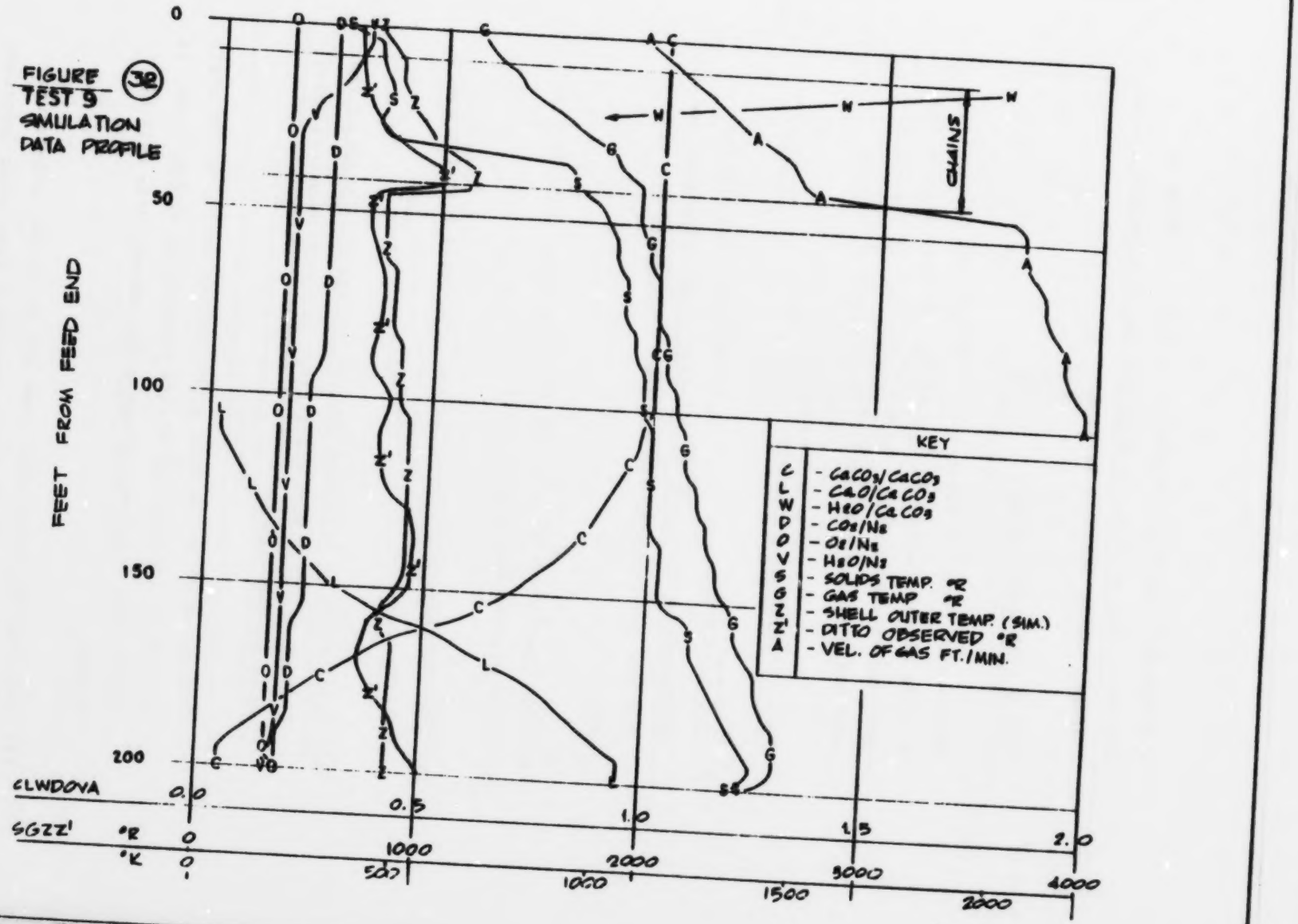

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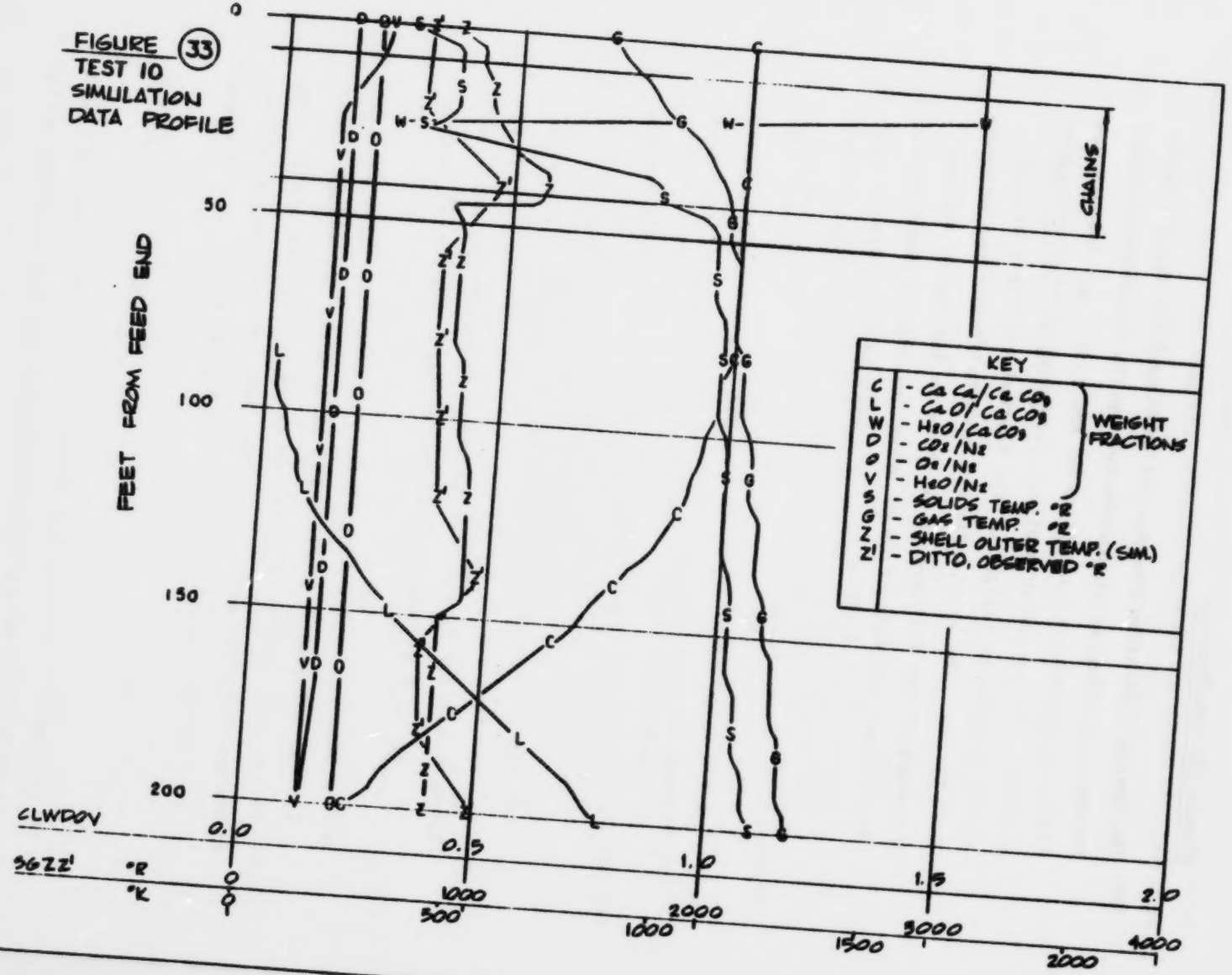
FIGURE (33)
TEST 10
SIMULATION
DATA PROFILE



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Simulation Conclusions

The Boeing Simulation Program was developed for fossil-fuel firing, and required adaptation for use in these Tests with regard to convective heat transfer and radial gas temperature profile. The parameter profiles developed in Figs. 20 and 26 to 33, provide an insight of the thermal processes of dual-fuel firing. The simulation was used to predict optimal performance of the Test system as shown in Section 8, and may now be used for evaluation of other dual-fuel-fired lime kilns.

8. PREDICTION OF OPTIMAL PERFORMANCE

The Test series was performed by varying oil substitution and replacement ratio only. The computer simulation based on these data was then run at 50, 75 and 100% substitution, to examine the effect of various operational parameters on performance.

Optimization Assumptions

Base values for the simulation were:

Kiln rotational speed	= 1.2 rpm
Kiln length/internal diam.	= 200:6
Wet-Cell gas temp.	= 1200° C. (2200° F.)
Lime mud feed	= Baseline level 76.3 tonne/d constant
Cooling water added to mud feed	= Constant level above baseline
Dust loss	= Baseline level amended per Section 6 (Page 73)
Ingress air	= Excess air ratio @ 53% for oil (See discussion)

Ingress air consists of leakage air, combustion air for the oil and cooling air for the gun. (Section 6.) The test results indicate the importance of reducing air leakage and design modifications to achieve this are detailed in Section 8.

Ideally, only the combustion air for the oil is required at the hood. Excess air ratio of 15-20% would be normal for the oil gun. Total air for the baseline (100% oil-fired) test was equivalent to 53% excess air. For the simulation, this ratio is assumed to apply at all levels of oil-firing with increased substitution.

Optimization Techniques

From the simulation results, curves were plotted of lime availability and kiln exit gas temperature against Wet-Cell input at

50, 75 and 100% oil substitution levels (Fig. 34(a) and (b) respectively). Replacement ratio, defined by Equation (26), Page was also plotted against Wet-Cell input at various substitution levels in Fig. 34(c).

From Fig. 34(a) Wet-Cell input values for each level of substitution at a minimum permissible availability of 90% were then superimposed on Fig. 34(c). Similarly, Wet-Cell inputs from Fig. 34(b) were also superimposed on (c) using an exit gas temperatuve limit of 4000° C.

For the chosen set of base values, Fig. 34(c) shows how the 90% availability limit defined the minimum replacement ratio for optimum kiln operation at various substitution levels. Maximum permissible substitution for this case was limited by the exit gas temperature.

These data were then incorporated in the simulated performance curve of Fig. 35, relating replacement ratio to % oil substitution. The same technique was then used to compare the individual impact on kiln performance of the following variables as shown in Figs. 36, 37 and 38:

--Rotational speed change
--Length/diameter ratio
--Gas temperature at inlet
--In addition, influence of chain section length was investigated and effect of ingress air by simulation was compared to actual test.

Influence of Variables of Performance

Rotational Speed Change

The simulation was run for a kiln rotational speed of 0.7 rpm and the results compared with the base case of 1.2 rpm (Fig. 36).

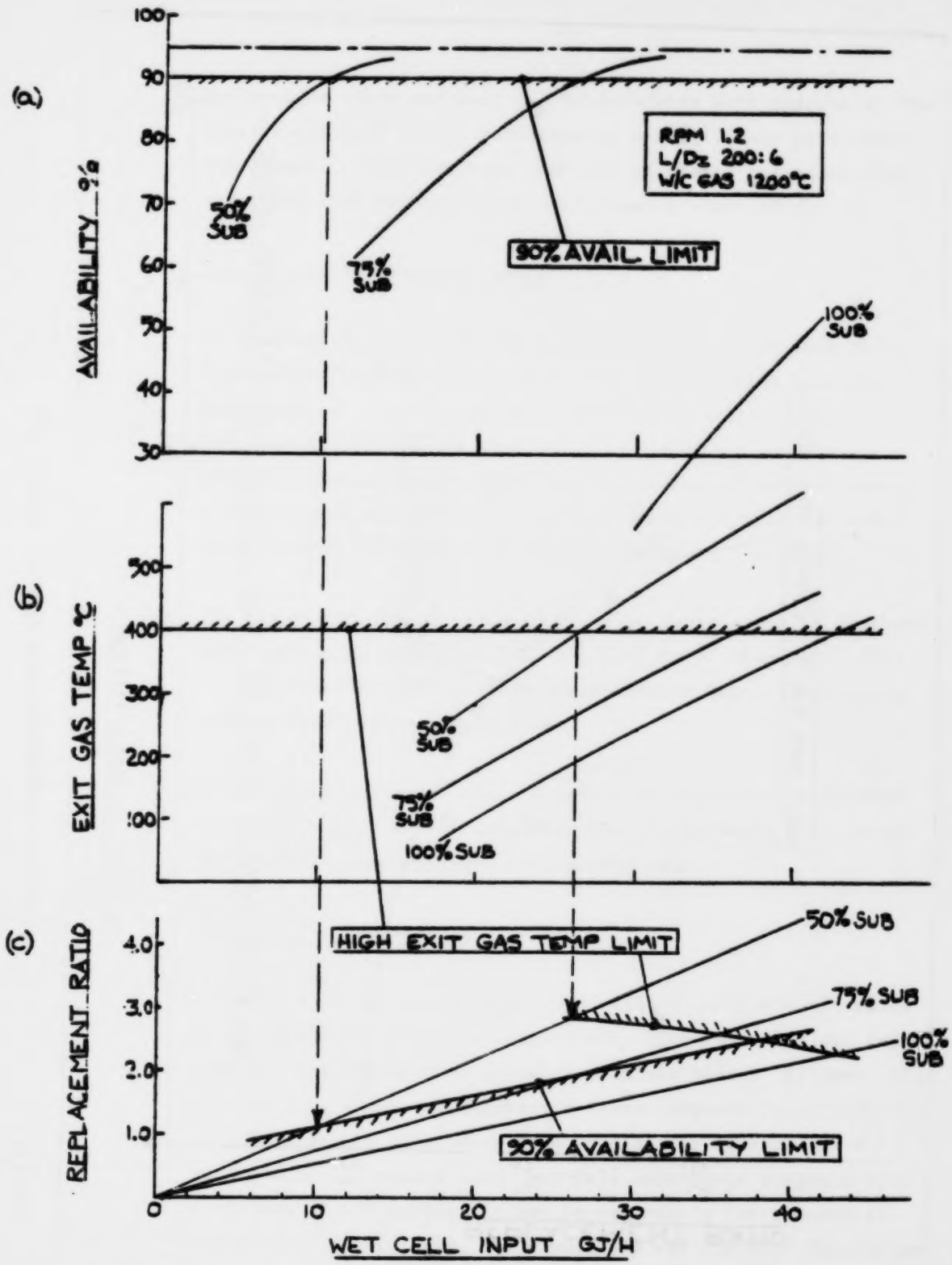

FIGURE 34 DEVELOPMENT OF PERFORMANCE LIMITS

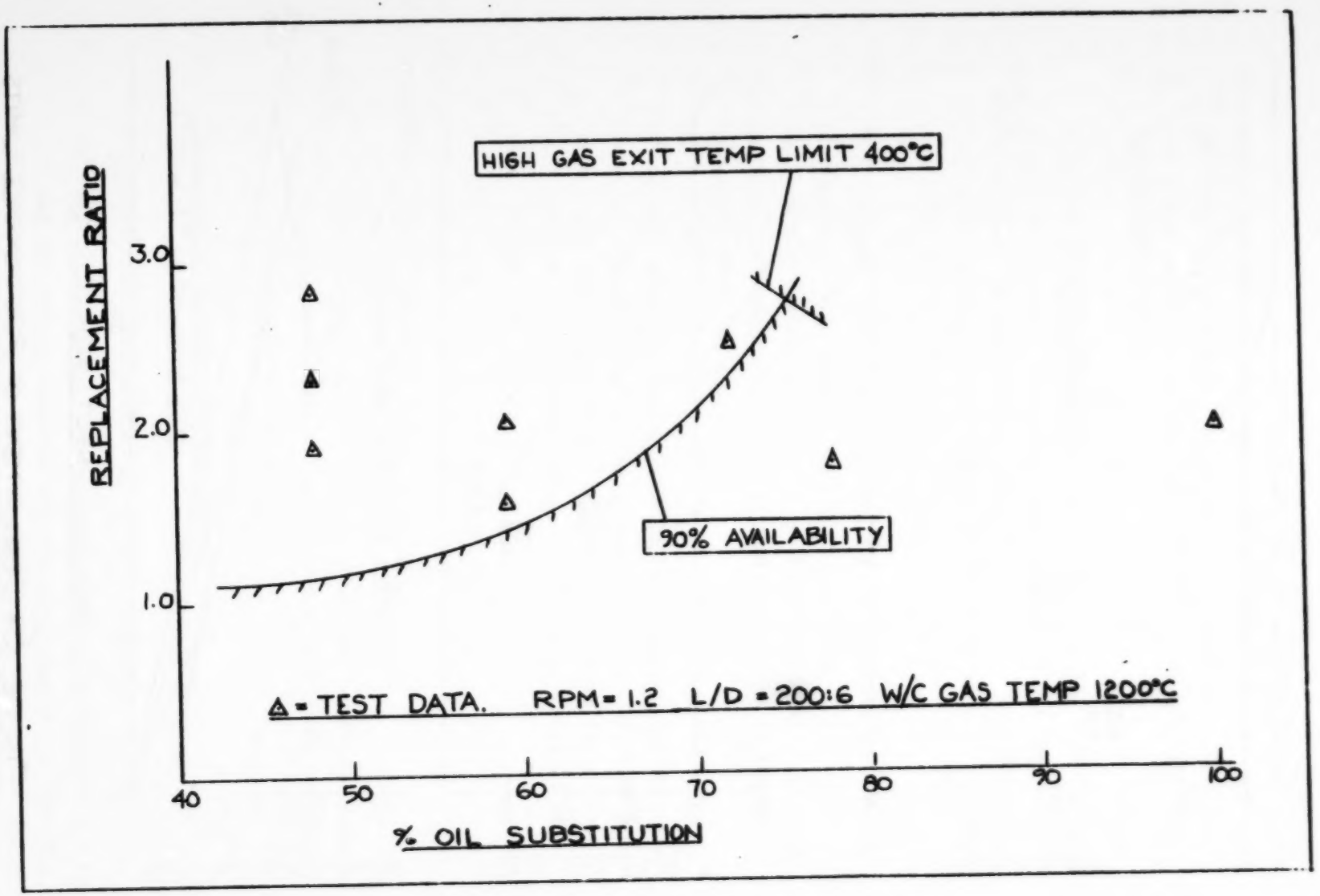

FIG. (35) SIMULATED PERFORMANCE WITH REDUCED INGRESS AIR

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Replacement ratio and exit gas temperatures were reduced at the lower rotational speed, representing a significant performance improvement. This confirms that the increased residence time resulting from reduced rotational speed is beneficial.

Length/Internal Diameter Ratio

An increase in length of the kiln would clearly improve performance by increasing residence time and reducing exit gas temperatures. In addition, reduction in internal diameter would increase gas mass velocity (kg/m^2-h) and thus enhance convective heat transfer. Reduction of shell diameter would require substantial structural changes, hence the computer simulation assumed increased refractory thickness.

The results of Fig. 37 show progressive improvement in replacement ratio with reduction from 6', 5.5' to 5' diameter. This is attributable both to enhanced convective heat transfer and reduced shell radiant heat loss.

In the practical case, these benefits of increased refractory thickness would entail consideration of increased kiln weight and possible acceleration of refractory wear.

Wet-Cell Gas Temperature

There is no doubt of the beneficial effect on kiln performance of increased Wet-Cell gas temperature. Theoretical flame temperatures are higher than those used in the tests. Tillman (5), quoting L.L. Anderson, suggests a flame temperature of 1500° C. (2732°F.) may be attainable from Douglas fir bark at 50% M.C. burned with 10% excess air. Wet-Cell experience suggests that a minimum of 15% excess air may be reasonable for commercial operation. Accordingly, the simulation was run with the following values:

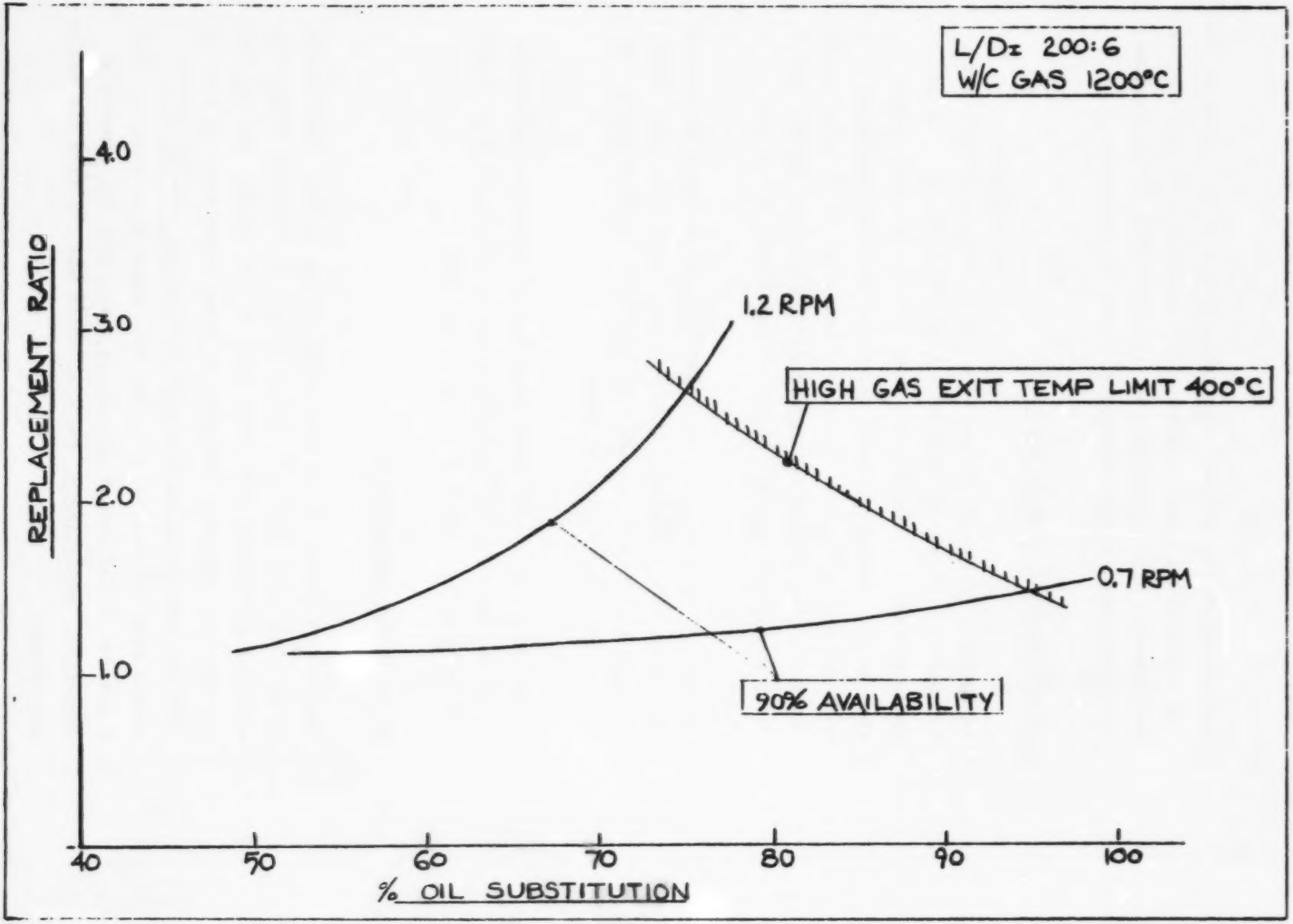

FIG. (36) SIMULATED PERFORMANCE WITH SPEED CHANGE

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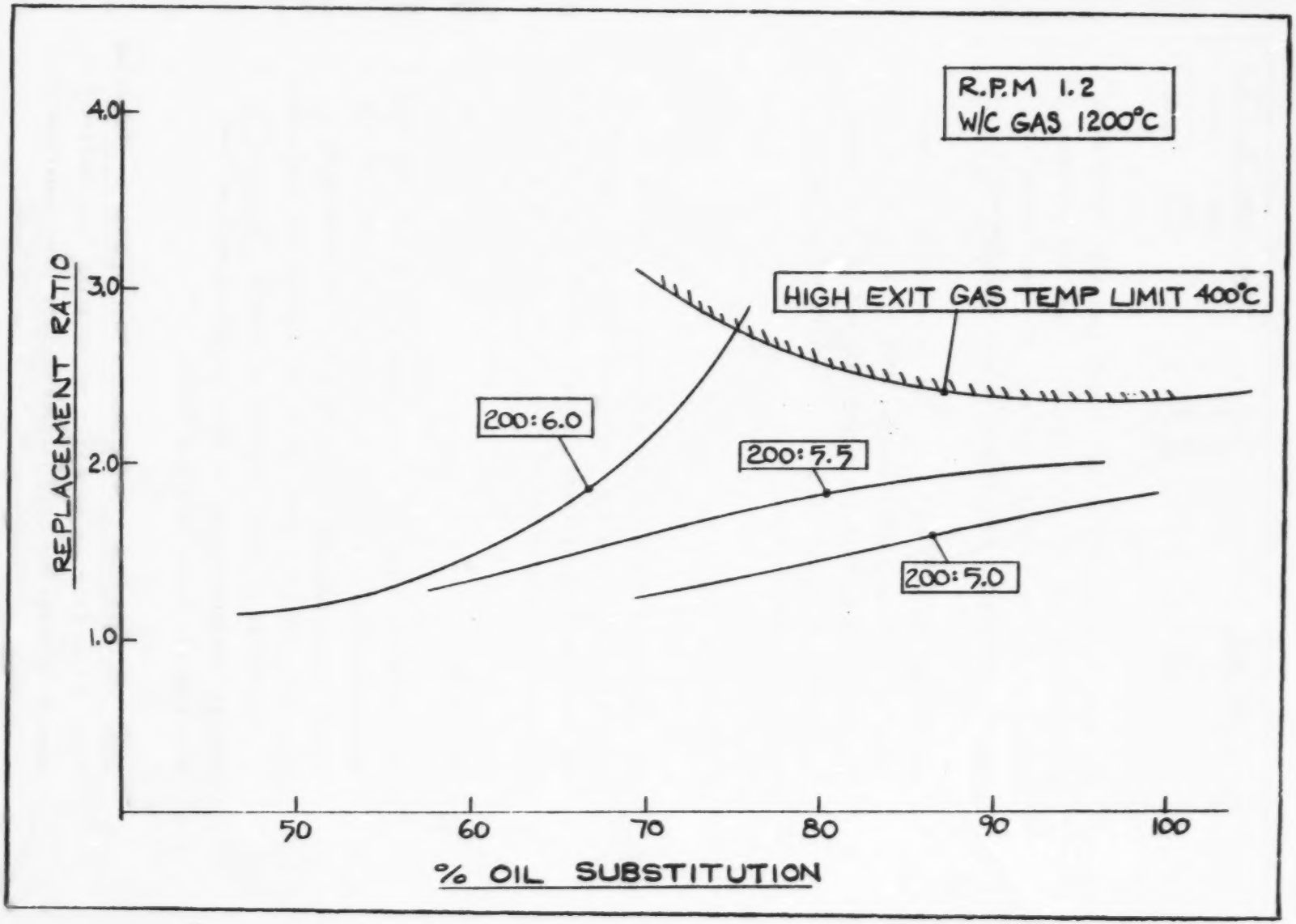

FIG. (37) SIMULATED PERFORMANCE WITH INSIDE DIAMETER CHANGE

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Fuel M.C. %	Max. Temp. °C. (°F.)
45	1404 (2560)
50	1324 (2415)
55	1232 (2250)

The simulation results (Fig. 38) show that kiln performance would improve substantially with drier fuel and increased gas temperatures. In addition, the exit gas temperatures are reduced, indicating more effective heat transfer.

For 50% and 45% M.C. fuel, exit gas temperatures no longer limit substitution, which reaches 100%. The significant improvement in performance between 55% and 50% M.C. suggests that some fuel drying may be economically attractive and worthy of investigation.

Chain Section

The simulation indicated that drying of the lime mud was achieved within the chain section for all cases examined. It was concluded that increase in chain section length would be of negligible benefit and no curves were developed.

Ingress Air

The simulation results for the assumptions of the base case were compared with the test data in Fig. 35. Test and simulated values converge at approximately 75% substitution. This indicates that reduction of air ingress has more impact on performance at lower substitution levels. This is because bulk gas temperature in the kiln is reduced more at lower substitution, i.e., lower gas flows.

It is concluded that close attention must be paid to minimizing tramp and cooling air in a kiln optimization (see Section 9), especially where economic factors dictate lower substitution.

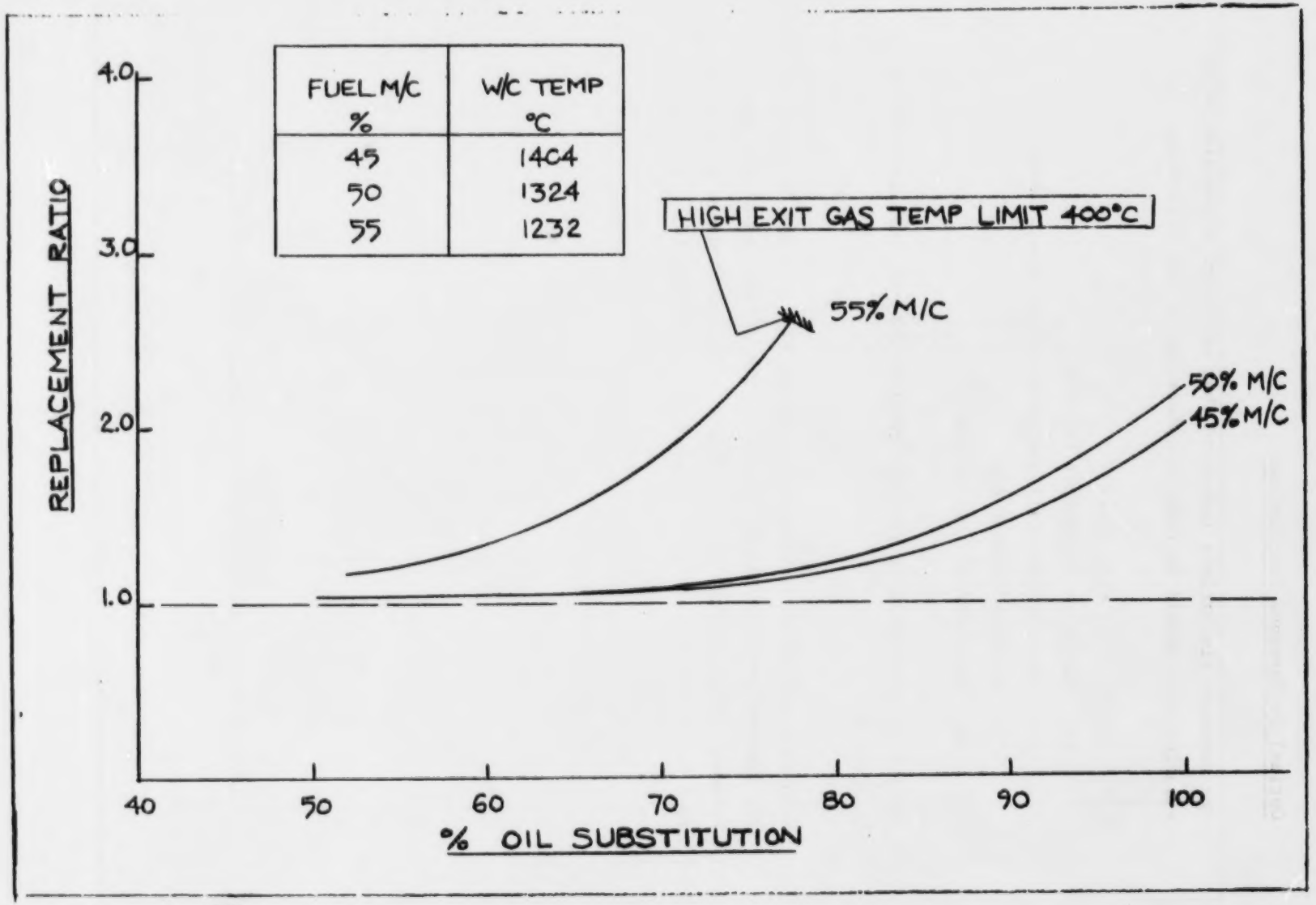

FIG. (38) SIMULATED PERFORMANCE AT ELEVATED THEORETICAL TEMPERATURES

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Optimal Performance Conclusions

The computer simulations indicated that efficient conversion of an existing kiln should include consideration of the following:

(a) Reduced kiln speed
(b) Increased refractory thickness
(c) Increased Wet-Cell gas temperature, i.e., lower Moisture Content fuel
(d) Reduction of ingress air

Practical approaches to achieve these objectives are discussed in Section 9.

Consideration of performances defined by Figs. 35 and 36 (reduced air ingress and kiln speed) permitted the definition of the performance shown on Fig. 39 at 1.0 RPM. This performance is used as a basis for the economic evaluation in Section 10.

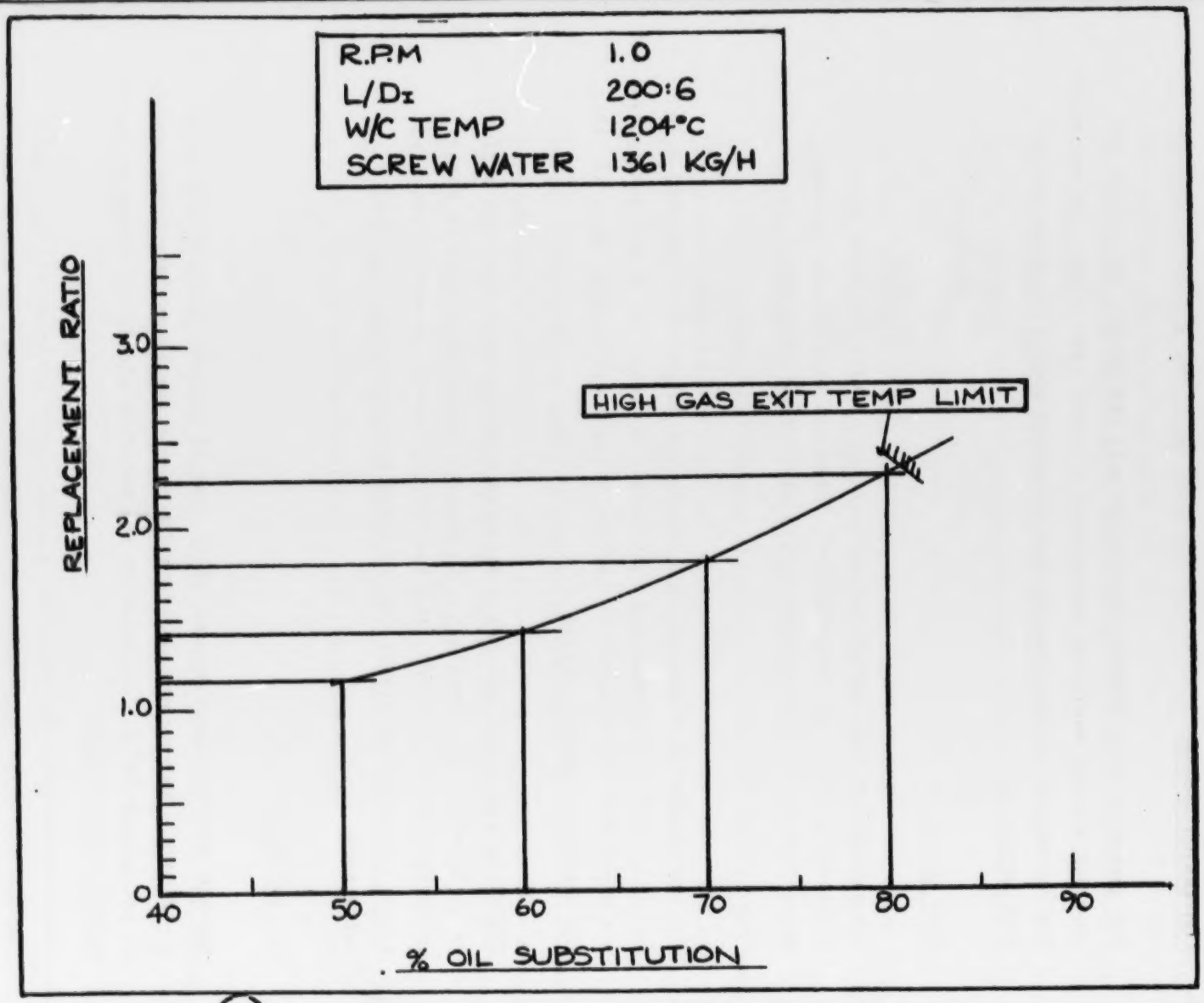

FIG. (39) SIMULATED PERFORMANCE FOR ECONOMIC EVALUATION

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9. CONVERSION DESIGN

The system at Port Alberni functioned well throughout the trials and provided a sound basis for conversion of other lime kilns. In approaching conversion design, special consideration should be given to the following areas:

Heat Source

The simulation highlighted the performance benefits arising from increased product gas temperatures. These can be achieved through Burner operation with reduced excess air and/or lower moisture content fuel.

The two-chamber Wet-Cell concept offers the potential to exploit both factors. The tests suggested that stable operation with minimum excess air might best be achieved by incorporating continuous O_2 analysis into the automatic control system.

Where the available fuel has a moisture content below 50%, the performance benefits are immediate. In coastal mills where fuel is wetter, investigation of site-specific fuel-drying economics may be required. Burner operation at higher temperatures would require improved refractory.

Hot Gas Duct

The hot gas duct constitutes a major loss of radiant heat, of the order 1 - 2 GJ/h. Refractory specification should ensure surface temperatures do not exceed 70° C. Location of the Wet-Cell should be chosen so as to minimize duct length.

Kiln Hood

Conceptual design of a kiln hood for dual firing is illustrated in Fig. 40. The design objectives were:

(a) to provide a smooth axial entry of Wet-Cell gases and promote oil flame stability.

(b) to reduce leakage air as far as possible.

(c) to eliminate direct impingement of the Wet-Cell gases on the kiln nose ring.

The duct is extended into the kiln, apart from a cutaway to permit lime drop-out. This almost completely eliminates hot gas impingement on the nose ring.

The oil gun assembly is extended through a 90° bend in the hot gas duct. The inside portion of the oil gun assembly is water-cooled. The primary (inner) air tube is equipped with spoilers or a turbulence shield per Fig. 10.

An additional cowling added to the short horizontal section of the gas duct provides support for the grate and a surface against which a seal ring can bear. The hood would be on wheels and equipped with clean-out doors and flame-monitors in the normal way.

The proposed design eliminates the need for cooling air on the nose ring and oil gun, providing substantial efficiency benefits.

Kiln

The converted kiln should be operated at reduced speed. The increased burden and additional load on the trunnions, will not normally be a problem.

Fan capacity required to accomodate increased gas flow after conversion will depend on substitution level and replacement ratio but can be expected to increase 30 - 100%.

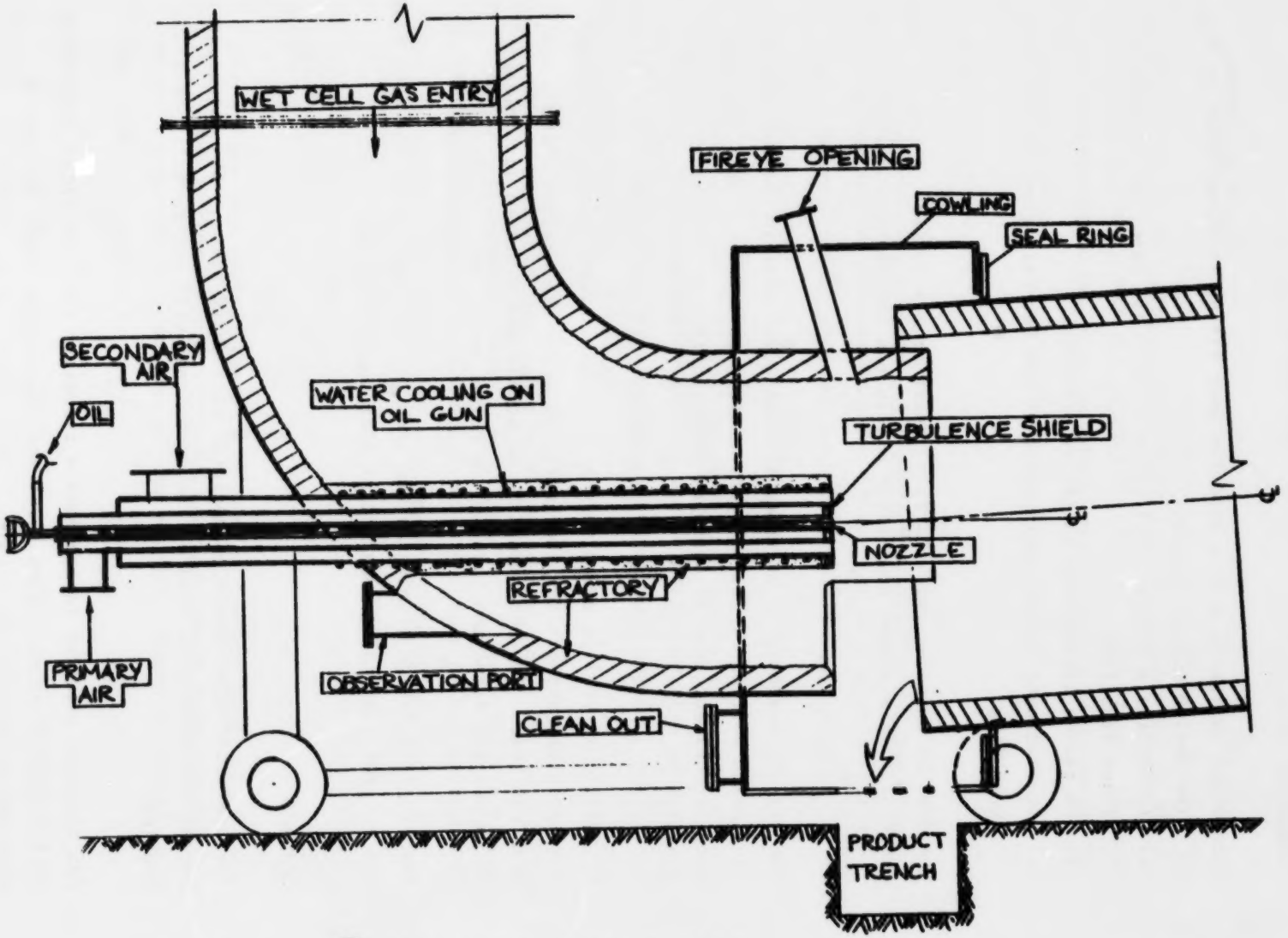


FIGURE (40) HOOD FOR DUAL FIRING

The fan should be located preferably after the dust scrubber since the gas temperature, fan wear due to dust load and volumetric capacity are lower than in the upstream location. Increased gas flow with dual-firing creates higher dust carryover. Conversion design must therefore reflect this in terms of dust loss and scrubber capability. It is recommended that the scrubber should be conservatively rated to handle 15% of dry solid feed after conversion. Stack loss on test increased from a maximum of 1.5% on baseline (oil-only) to higher levels on dual-firing. Where compliance regulations are stringent, scrubber performance must be improved to limit this.

It should be noted that implementation of the previously suggested measures (slower kiln rotation; reduced ingress air; increased gas temperature) would reduce scrubber and fan duty, bringing them closer to oil-only levels.

10. ECONOMIC ANALYSIS OF KILN CONVERSION

Introduction

The evaluation of the test series and computer simulation identified the principal engineering factors which should be considered for optimization of a lime kiln conversion. The choice of conversion approach, however, will be dictated generally by economic factors. These are analyzed in the discussion which follows.

Method

Net present value at 21% discount and internal rate of return were calculated for a lime kiln rated at 100 MM Btu/h on No. 6 oil fuel, operating at 90% capacity, with wood heat substitution of 50 - 75%. Hog fuel costs ranged between -$5/unit and $30/unit, with oil costs $20/bbl to $40/bbl. Two different capital cost allowance schemes were applied, i.e.,

(a) straight-line depreciation over ten years, and
(b) accelerated two-year depreciation.

Scheme (b) is available in Canada for equipment that replaces or conserves fossil fuels. The calculation assumes a combined Federal and Provincial tax rate of 50% on project net cash flows minus the capital cost allowance.

In all cases the project is assumed to have an installation time of one year, an economic life of ten years, and zero scrap value. This means that all capital costs occur in Year 0 and that the project operates during Years 1 through 10. Fuel and operating and maintenance costs are assumed to escalate 10% annually due to inflation.

Capital Costs

The capital cost estimates for the system include Wet-Cell Burner, refractory-lined duct, kiln modifications and system installation. (Fig. 41). Larger wood burners show economies of scale up to 75-MM Btu/h output. For sizes (not yet designed) above 75-MM Btu/h, it was assumed that unit costs are constant.

The cost estimates are for standard units excluding research and development effort which may be required on certain applications.

Operating Costs

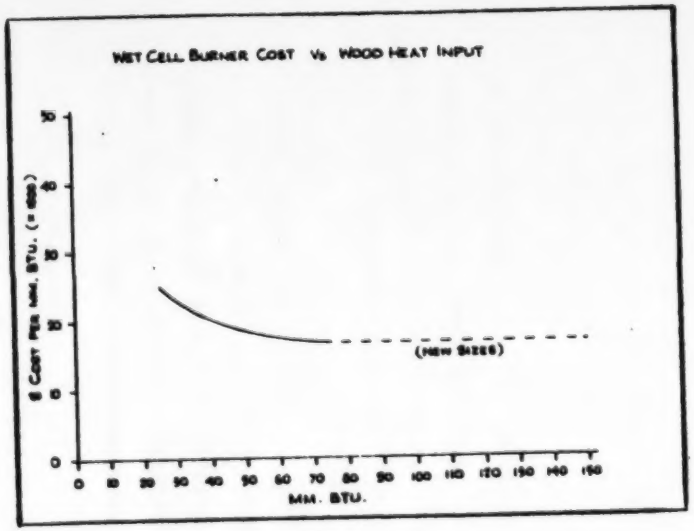
Operating costs consist of electrical, propane and maintenance. (Fig. 42). Annual maintenance costs were assumed to be 2% of system capital cost, with electrical and propane costs at 3¢/kwh and 21¢/litre, respectively. Propane costs are quite insignificant, as propane is used only for start-up. Total annual operating costs amounted to 5-7% of capital cost.

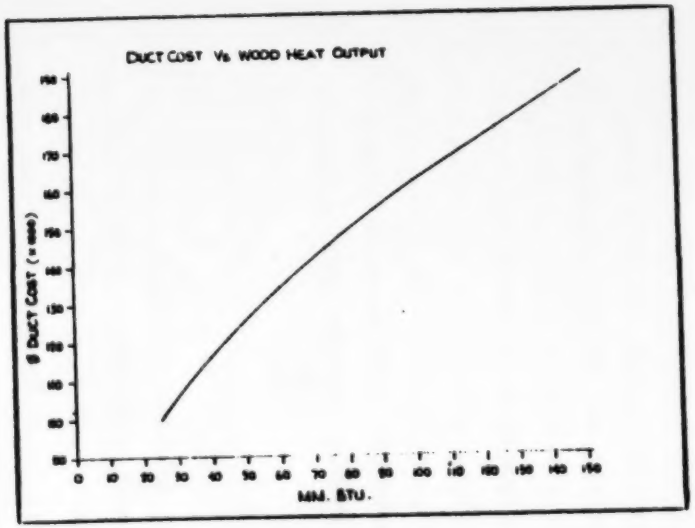
Fuel Costs

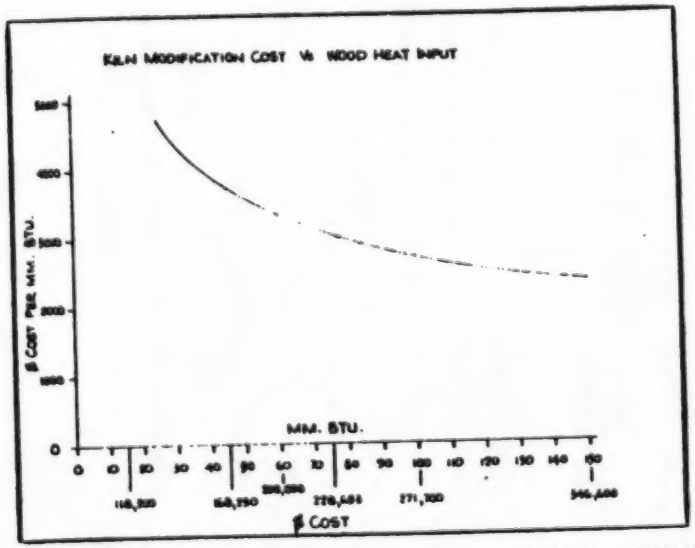
Hog fuel was assumed to have a higher heating value of 8500 Btu/BD lb and 55% moisture content (wet basis). No. 6 oil has a heating value of 6.27-MM Btu/bbl. Economic analyses were performed for hog fuel costs ranging from -$5/unit to $30/unit[1] and oil prices of $20, $30 and $40/bbl, in 1982 dollars.

Typical hog fuel costs in the Port Alberni region are toward the higher end of the modelled range. Hog fuel costs are negative when disposal expense is incurred. Such a situation can occur where there is a large surplus of wood waste, as in central British Columbia.

(1) One unit = 200 cu ft = 4440 wet lbs @ 55% M.C.
One unit = 9.439 m^3 = 2014 wet kg @ 55% M.C.







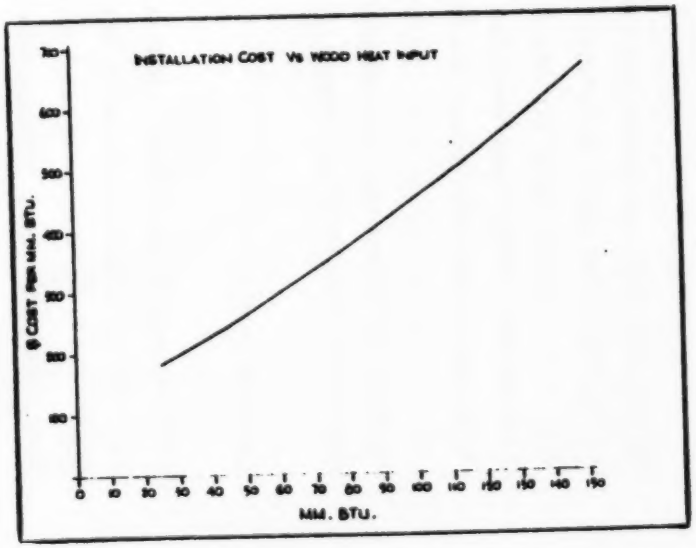
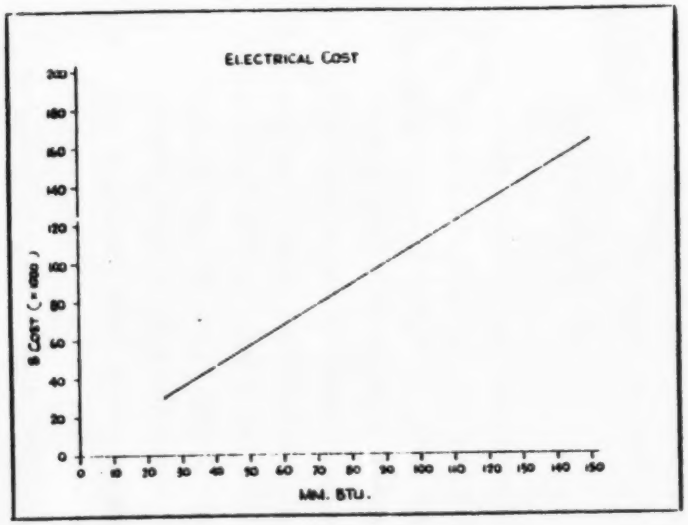
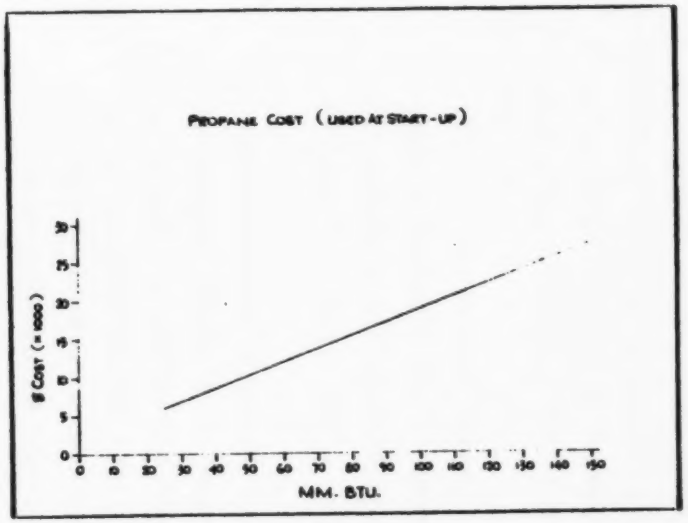


FIG. (41) CAPITAL COSTS VS. WOOD HEAT INPUT





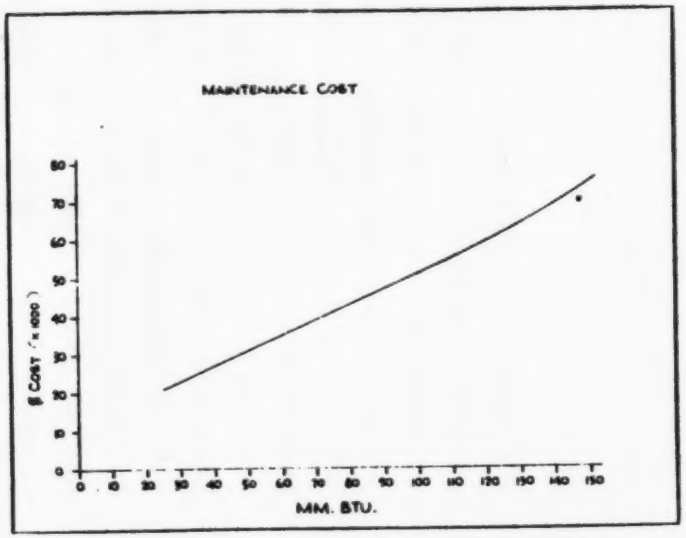


FIG. (42) **OPERATING COSTS VS. WOOD HEAT INPUT**

The analytical method is illustrated for a base case of $10/unit hog fuel and $30/bbl oil.

Evaluation Criteria

Two evaluation criteria were used -- internal rate of return (IRR) and net present value. Net present value is now universally considered to be the preferred criterion in cost/benefit analysis. The two criteria are briefly discussed.

The net present value of the project is the sum of the discounted net cash flows in each year. The discount rate of 21% consists of a 10% inflation rate and a 10% real cost of capital (1.1 x 1.1 = 1.21). Present values are in 1982 dollars. The present value criterion asserts that the project is worthy of investment when its net present value is greater than zero. The net present value is simply the profit of the project, with future costs and incomes adjusted for inflation and the fact that money received or spent in the future is worth less than money received or spent now, i.e., the real cost of capital.

The internal rate of return is the discount rate which makes the present value equal to zero. The IRR criterion is that the project is worthy of investment when the IRR exceeds the rate of return that could be made on an alternative investment (for example, the interest rate). Problems with using IRR as an investment criterion have been discussed, e.g., (12). For example, a two-year project with net cash flows of -1 dollar and 3 dollars has a higher rate of return (200%) than a project with net cash flows of -1,000 dollars and 2,500 dollars (150%), but clearly the second project is preferable to the first. The present value of the second project (using a 21% discount rate) is $881.00, whereas the present value of the first is only $1.22. Consequently, one project may show a very much higher IRR than another, when it is clearly inferior.

The internal rate of return may not even be unique. There may be several rates of return which make the present value of the project equal to zero. This will frequently occur in projects with uneven cash flows -- where benefits may exceed costs in some years but not all years.

However, despite the above inadequacies, the internal rate of return method is frequently used for evaluation and is therefore provided for completeness.

A third criterion sometimes used in project evaluation is the payback period. This is the number of years from the beginning of the project until the time when accumulated net cash flows first exceed the capital cost. This calculation ignores all costs and benefits beyond the payback period and provides no criterion for investment. It has the advantage of simplicity, but ignores the cost of capital (discounting). It is inferior to even the least rigorous of present value calculations.

Analysis of Results

The present values of net benefits for various oil and hog fuel prices are presented in Table 13. This table assumes that the project is written off on a straight-line basis for tax purposes. Table 14 shows the effect of the accelerated capital cost allowance on the net present value of the project, for a fixed pair of hog fuel and oil prices. Tables 15 and 16 show annual cash flows for the two tax cases and Table 17 shows the internal rate of return for various oil and hog fuel prices with two different levels of oil substitution. Table 17 assumes straight-line write-off. Table 18 compares internal rates of return for the two tax cases.

TABLE 13 NET PRESENT VALUE FOR VARIOUS OIL AND HOG FUEL COSTS

(000 1982 $) (Based on Figs. 41 and 42)

HOG FUEL COST ($/unit)	OIL COST ($/bbl)											
	20				30				40			
	Oil Substitution Level				Oil Substitution Level				Oil Substitution Level			
	50%	60%	70%	75%	50%	60%	70%	75%	50%	60%	70%	75%
-5	1956	2245	2147	2300 *	3385	3958	4375	4442 *	4814	5673	6376	6584 *
0	1553	1654 *	1503	1230	2982	3367	3500 *	3372	4411	5082	5501	5515 *
10	748 *	472	(215)	(909)	2177	2208 *	1750	1230	3605	3900 *	3752	3375
20	(58)	(710)	(1023)	(3054)	1371 *	1049	1	(907)	2800 *	2718	2002	1230
30	(863)	(1182)	(2361)	(5196)	565 *	(110)	(1749)	(3047)	1994 *	1536	245	(912)

(*) The optimum economic level of substitution

TABLE 14

EFFECT OF CHANGE IN CAPITAL COST ALLOWANCE

ON NET PRESENT VALUE

($000)

(Oil $30/bbl -- Hog Fuel $10/unit)

	Substitution Levels		
	50%	60%	75%
Straight-Line Depreciation	2177	2208	1230
Two-Year Depreciation	2400	2602	1891

TABLE 15

CASH FLOW STATEMENT ($000)

FOR OPTIMUM LEVEL OF SUBSTITUTION

Hog Fuel ($10) -- Oil ($30)

Substitution Level (60%)

Straight Line Depreciation

	0	1	2	3	4	5	6	7	8	9	10
SAVINGS--BUNKER C											
Current Oil Cost		3,712	4,083	4,492	4,941	5,435	5,978	6,576	7,234	7,957	8,753
SAVINGS		2,227	2,450	2,695	2,964	3,261	3,578	3,946	4,340	4,774	5,252
COSTS											
Capital Investment	2,200										
Hog Fuel		510	561	617	679	747	821	904	994	1,093	1,203
Oper. & Maint.		162	178	196	216	237	261	287	316	347	382
TOTAL COST		672	739	813	894	984	1,082	1,190	1,310	1,440	1,585
CASH FLOW (Before Taxes)	(2,200)	1,555	1,711	1,882	2,070	2,277	2,505	2,755	3,031	3,334	3,667
TAXES											
Depreciation	200	200	200	200	200	200	200	200	200	200	200
Net Tax Value (50%)	100	(678)	(755)	(841)	(935)	(1,038)	(1,152)	(1,278)	(1,415)	(1,567)	(1,734)
CASH FLOW (After Taxes)	(2,100)	878	955	1,041	1,135	1,238	1,352	1,478	1,615	1,767	1,934

LAMB - CARGATE INDUSTRIES LIMITED

TABLE 16

CASH FLOW STATEMENT ($000)

FOR OPTIMUM LEVEL OF SUBSTITUTION

Hog Fuel ($10) -- Oil ($30)

Substitution Level (60%)

Two-Year Depreciation

	0	1	2	3	4	5	6	7	8	9	10
SAVINGS--BUNKER C											
Current Oil Cost		3,712	4,083	4,492	4,941	5,435	5,978	6,576	7,234	7,957	8,753
SAVINGS		2,227	2,450	2,695	2,964	3,261	3,587	3,946	4,340	4,774	5,252
COSTS											
Capital Investment	2,200										
Hog Fuel		510	561	617	679	747	821	904	994	1,093	1,203
Oper. & Maint.		162	178	196	216	237	261	287	316	347	382
TOTAL COST		672	739	813	894	984	1,082	1,190	1,310	1,440	1,585
CASH FLOW (Before Taxes)	(2,200)	1,555	1,711	1,882	2,070	2,277	2,505	2,755	3,031	3,334	3,667
TAXES											
Depreciation	1,100	1,100									
Net Tax Value (50%)	550	(228)	(855)	(941)	(1,035)	(1,138)	(1,252)	(1,378)	(1,515)	(1,667)	(1,605)
CASH FLOW (After Taxes)	(1,650)	1,328	855	941	1,035	1,138	1,252	1,378	1,515	1,667	1,605

LAMB-CARGATE INDUSTRIES LIMITED

TABLE 17

INTERNAL RATE OF RETURN FOR VARIOUS OIL AND HOG FUEL COSTS AT 50% AND 75% SUBSTITUTION

HOG FUEL COST ($/unit)	OIL COST ($/bbl)					
	20		30		40	
	Oil Substitution Level		Oil Substitution Level		Oil Substitution Level	
	50%	75%	50%	75%	50%	75%
-5	54	39	74	53	94	67
0	48	31	69	46	89	60
10	35	12	56	31	75	46
20	20	Neg	45	12	66	31
30	17	Neg	32	Neg	54	12

TABLE 18

EFFECT OF CHANGE IN CAPITAL COST ALLOWANCE ON INTERNAL RATE OF RETURN

Oil $30/bbl -- Hog Fuel $10/unit)

	Substitution Levels		
	50%	60%	75%
Straight-Line Depreciation	56	49	31
Two-Year Depreciation	77	68	43

The most interesting aspect of Table 13 is the variation in optimum substitution level with different combinations of hog fuel and oil costs. This arises from the fact that increase in substitution level (Fig. 40) causes requirements for wood heat to increase much faster than proportionally. For this reason, although system capital costs show economies of scale with respect to wood heat output, no economies of scale occur with respect to quantity of oil replaced. For example, increasing oil substitution from 50% to 75%, increases burner size required from 59 to 155 MM Btu/h and capital cost of the project from $1.65-million to $3.8-million; i.e., more than double.

For the same reason, increased substitution requires a more than proportional increase in hog fuel needs. For a shift from 50 to 75% oil subsitution, a 50% increase in oil savings, hog fuel costs will rise 76%.

In summary, the capital cost and hog fuel cost both increase more than proportionally for a given increase in oil substitution. However, the net profitability of the project and hence the optimum substitution level, will depend on the unit cost of the oil replaced.

The calculations of net present values of kiln conversion projects ranging from 50% to 75% oil substitution show that the optimal level of conversion may lie anywhere between 50% and 75%, depending on the prices of oil and hog fuel. As long as oil is very expensive, or hog fuel is very cheap (or has a negative cost) it will be best to replace as much oil as possible. Similarly, cheap oil or expensive hog fuel means that if any project is economic, it will be the one replacing the least oil. For intermediate hog fuel and oil prices, the optimal level of substitution will be quite sensitive to changes in fuel prices.

Altering the tax scheme from a straight-line to a two-year write-off (Table 14) increases the project's profitability more as the level of

oil substitution rises. This is because of the higher investments required by higher levels of substitution. Consequently, it is possible that, with a two-year write-off, a high substitution level could be preferred to a low one where, with straight-line depreciation, the opposite would be true. Thus, it is likely that each firm will have to apply its own tax rates and capital cost allowance schemes to the project analysis in order to determine the best level of substitution, even if hog fuel and oil prices are known with certainty.

A final and important consideration in the choice of the best substitution level is the moisture content of the hog fuel. The calculations presented here assume that the hog fuel is 55% moisture content (wet basis). A drop in the moisture level will increase the efficiency of wood burning, effectively lowering hog fuel costs (as long as these are positive) and raising the optimal substitution level. A drop in moisture content from 55% to 45% would lower total hog fuel costs by about 30%, possibly increasing the best level of substitution for hog fuel in the $0 to $20 range. A further 10% drop in moisture content to 35% would cut hog fuel costs another 20%.

Conclusions

1. Based on current world residual oil price, lime kiln conversion is economic, if hog fuel is available for $30/unit or less.

2. The optimal level of oil substitution will vary with the prices of oil and wood waste, the capital cost allowance scheme in effect, and the moisture content of the wood waste. For low oil or high hog fuel prices, a low substitution level (50%) is preferred. For high oil prices or low hog fuel prices, a high level (75%) is best. For intermediate fuel prices, an intermediate level of substitution is preferable.

3. The introduction of a two-year tax write-off substantially improves the project's profitability and has a greater effect on profits for projects replacing more oil. Shifting from straight-line depreciation to a two-year write-off will, in some instances, make higher substitution levels preferable.

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STANDARD: 523

DATE EFFECTIVE: May 25, 1965

REFERENCE PROCEDURE:

AVAILABLE CaO IN REBURNED LIME

Object: To determine the available calcium oxide in quicklime or hydrated lime.

Equipment:
(1) Mortar and pestle
(2) 100 mesh screen
(3) 250 ml Erlenmeyer flask
(4) 2000 ml Erlenmeyer flask
(5) 50 ml pipette
(6) Hot plate
(7) Magnetic stirrer
(8) Watch glass
(9) 50 ml burette
(10) Balance sensitive to 1 milligram

Reagents:
(1) 1.0 N HCl
(2) Phenolphthalein indicator
(3) CO_2 free distilled water - (boil approximately 1 liter of distilled water just previous to doing test)
(4) Granulated sugar (sucrose)

Procedure:

(1) Obtain a representative portion of the lot to be analysed and after mixing well, weigh out approximately 150 grams.

(2) Grind sample with mortar and pestle until it all passes the 100 mesh screen. (Do not discard anything) Mix well.

(3) Weigh exactly 0.500 grams of prepared sample on a previously tarred watch glass.

(4) Add 50 mls of CO_2 free water to a 250 ml Erlenmeyer flask and then brush the weighed sample into the flask.

(5) Immediately stopper the flask and disperse the sample thoroughly by swirling.

(6) Bring the sample to the boil for 3 minutes with the bottle loosely stoppered.

(7) Remove the stopper and add 100 mls of CO_2 free water.

(8) Add 15 grams of granulated sugar (sucrose). Shake vigorously at intervals over a 5 minute period.

(9) Let sample stand for at least 30 minutes but less than one hour.

(10) Remove stopper, drop bar magnet into flask and place on magnetic stirrer.

(11) Add three drops of phenolphthalein indicator.

(12) Titrate with 1.0N HCl to disappearance of pink.

Calculation:

$$\% \text{ CaO} = 5.60 \times \text{mls } 1.0\text{N HCl}$$

Factor Derivation:

A 0.5 gram sample of reburned lime and 1.0 N HCl are used.

$$\text{gms CaO} = \frac{\text{Titration} \times 1.0 \times 36.5 \times 56}{1000 \qquad 73}$$

$$\% \text{ Available CaO} = \frac{\text{gms CaO} \times 100}{\text{Wt. of sample}}$$

$$= \frac{\text{Titration} \times 1.0 \times 36.5 \times 56 \times 100}{1000 \times 0.5 \times 73}$$

$$= \text{Titration} \times 5.60$$

Reaction: Reburned lime is slaked to form $Ca(OH)_2$. The $Ca(OH)_2$ is then titrated with HCl. Sugar is added to help keep $Ca(OH)_2$ in solution and to precipitate out $Mg(OH)_2$.

Prepared By: E D Cooling Lam

Approved:

Columbia Cellulose Company, Limited
Celgar Pulp Division

SCANDINAVIAN PULP, PAPER AND BOARD
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SCAN-N 25:81

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Burnt lime sludge

Available Lime

1. Scope and field of application

This SCAN-test Standard specifies a method for the determination of available lime in burnt (reburnt) lime sludge. The procedure is based on that specified in ASTM Standard C-25 and should give equivalent results.

2. Definition

Available lime: An arbitrary measure, describing the content of such calcium oxide that enters into reaction under the conditions specified in this Standard. The result is expressed as a percentage of the mass of the sample as taken.

NOTE – Magnesium oxide may be present in lime. Under the conditions of this test only a few percent thereof enters into reaction.

3. Principle

The burnt lime sludge is ground to a fine powder and slaked with distilled water. Sucrose is added to accelerate the dissolution of calcium hydroxide. The resulting alkaline solution is titrated with hydrochloric acid and the amount of acid consumed is used to calculate the percentage of available lime.

4. Reagents

4.1 *Carbon dioxide-free distilled water.*
Prepare by boiling distilled water. Allow to cool in a stoppered bottle. Prepare a new batch each day samples are analysed.

4.2 *Sucrose*, $C_{12}H_{22}O_{11}$
Sucrose (sugar) in granular form for domestic use is normally acceptable. Check for acidity by titrating an aqueous solution, made from 15 g of the sugar, with 0.1 mol/l sodium hydroxide solution, using phenolphthalein as the indicator. If more than 0.1 ml are required to produce a pink colour, the sugar contains acids and should not be used for this test.

4.3 *Hydrochloric acid*, standardized *ca* 1 mol/l
The concentration should be known to the nearest 0.001 mol/l.

4.4 *Phenolphthalein indicator solution*, 0.1%.
Dissolve 0.1 g of phenolphthalein in 100 ml of ethanol, C_2H_5OH.

5. Preparation of sample

Samples should be protected from exposure to air and should be analyzed with a minimum of delay.

Break up any lumps an obtain a representative sample of about 75 g. Grind this portion in a mill or equivalent, avoiding undue exposure to air, until it passes a sieve with a nominal sieve opening of 150 μm. Then thoroughly mix the ground material and place it in an air-tight container.

6. Procedure

Carry out this procedure with duplicate samples.

Pour about 10 ml of distilled water, free from carbon dioxide, in a 300 ml conical flask.

Weigh a 0.5 g portion of prepared sample to the nearest 1 mg and carefully brush it into the flask. Stopper the flask immediately.

Disperse the sample thoroughly in the water by a swirling motion.

Remove the stopper and place the flask on a hot plate and immediately add 150 ml of boiling carbon dioxide-free water. Swirl the flask and boil actively for 1 min to achieve complete slaking.

Remove the flask from the hot plate, stopper it loosely and place it in a cold water bath to cool it to room temperature.

Add 15 g of sucrose, stopper the flask, swirl, and allow it to stand for 15 min to react. Swirl at 5 min intervals during the reaction period. Alternatively a magnetic stirrer may be used.

NOTE. – The amounts of water and sugar to be used are only approximate. To save time, a small measure can be used, which when full holds 15 g of the sugar used.

Remove the stopper, add 4 to 5 drops of the phenolphthalein indicator solution. Wash down the stopper and the sides of the flask with carbon dioxide-free water.

Titrate rapidly with the standardized hydrochloric acid. Swirl the flask or stir the solution during the entire titration. When the first complete disappearance of the pink colour is observed, read the end point. Ignore the return of the pink colour.

NOTE. – Unless the operator is familiar with a previous analysis of the lime under test, it is good practice to carry out a preliminary titration.

7. Calculation

Calculate the percentage of available calcium oxide from the expression

$$X = 2.8\,ac/w$$

where

a = volume of hydrochloric acid consumed in the titration, millilitres.

c = concentration of hydrochloric acid, moles per litre.

w = amount of sample, analysed as taken, grams

X = the content of available calcium oxide, per cent.

The numerical factor contains the relative molar mass of 0.5 moles of calcium oxide, 28.0, and the factor 0.1 necessary to convert the result to percentage units.

8. Precison

A sample of burnt lime sludge was analysed by this procedure by 8 laboratories. The mean value obtained was 84.5 per cent. None of the results was excluded. The relative standard deviation (coefficient of variation) was 1.0 per cent. The 95 per cent confidence limits were ± 2.0 per cent.

9. Report

Report the result as a percentage to the first decimal place. Report each result separately.

The test report should include a reference to this Standard and the following particulars:

(a) date and place for testing
(b) identification mark of the sample analysed
(c) the results
(d) any departure from the standard procedure or any other circumstances that may have affected the results.

SCAN testing standards are issued and recommended by the Central Laboratories of the Pulp, Paper and Board Industries in Denmark, Finland, Norway and Sweden. Distribution: Secretariat, Scandinavian Pulp, Paper and Board Testing Committee, Box 5604, S-114 86 Stockholm, Sweden.